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Owens-Corning Fiberglas Corp, Toledo, Ohio

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The report details the preliminary engineering work done at Owens-Corning's (O-C's) Anderson, South Carolina, fibrous glass plant. The purpose of the work was to test, on a pilot plant scale, various technologies to be used to clean up industrial wastewater for a closed-loop system; i.e., for total industrial wastewater reuse. Conceptual design has been developed for the testing treatment processes of sand filtration, activated carbon adsorption, and disinfection. As a result of this work, O-C has authorized the construction of a full scale plant which will be in operation in 1978. This report makes the developed technology available to the industry prior to publication of details of final plant construction and operation.

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by

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CHAPTER I

CONCLUSIONS AND RECOMMENDATIONS

CONCLUSIONS

Analyses of accumulated data and the results of analytical and process investigations resulted in the following conclusions:

1. Reuse of treated industrial wastewaters in the process areas and cooling systems at the Anderson Plant is technologically feasible as proven in pilot cooling studies.

2. Reclamation of the effluent from the existing wastewater treatment facilities by sand filtration, carbon adsorption, and disinfection will result in a product water suitable for reuse.

3. Projected industrial wastewater flows and reuse requirements for a recirculation system at the Anderson Plant indicate that during the summer season reuse will exceed wastewater production by 54.3 gpm; conversely, during the winter months, production will exceed reuse by approximately 6.4 gpm.

4. Based upon current drift loss estimates, equilibrium concentrations of total hardness, calcium hardness, silica, sulfate, and zinc in the cooling systems may exceed the water quality criteria for these uses if drift loss is the sole mechanism by which inorganic dissolved solids are removed from the reclaimed wastewater.

5. If removal of inorganic dissolved solids is required, it will be accomplished through treatment by reverse osmosis or lime-soda softening/anion exchange.

6. Performance data for the existing wastewater treatment facility exhibit exceptional BOD_5, COD, TOC, and TSS removal efficiencies.

7. Coagulation of equalized raw wastewater with combinations of ferric chloride, clay, and cationic polymer results in improved TSS removal efficiencies as observed during bench scale laboratory tests and full scale operations.

8. Dissolved air flotation without chemical addition is an effective pretreatment process for removal of fibrous materials from Mat Line wastewater. However, laboratory tests performed with combined raw wastewaters obtained equivalent treatment performances for both dissolved air flotation and sedimentation.

9. Complete physical-chemical treatment of Anderson Plant raw wastewater is not feasible due to extreme residual concentrations of non-adsorbable organic compounds.

10. Carbon adsorption of Anderson biologically-treated effluent will produce a reclaimed wastewater which is suitable for plant reuse in terms of dissolved organic constituents.

11. Carbon exhausted through treatment of biological effluent is quite amenable to regeneration.

12. Both ozonation and chlorination will eliminate virtually all fecal coliform organisms from combined industrial-sanitary biologically-treated effluent.

13. The capacity of mixed bed ion exchange resin with respect to inorganic dissolved solids removal is quite low due to rapid breakthrough of silica.

14. Projected design loadings for advanced wastewater treatment processes are as follows:

    Flow = 285 gpm
    TSS Loading = 31 lb/day
    TOC Loading = 116 lb/day

RECOMMENDATIONS

Evaluation of all preliminary engineering efforts and the preceeding conclusions lead to the following recommendations:

1. Implement an industrial wastewater recirculation system at the Anderson Plant which includes the elements listed below.

   . Tertiary treatment of the effluent from the existing wastewater treatment facilities through sand filtration, activated carbon adsorption, and disinfection

- Segregation and separate treatment of all sanitary wastewaters
- Use of reclaimed wastewater in the process areas and cooling systems

2. Tertiary treatment facilities should conform to the design criteria summarized below:

Filter Feed and Backwash Sump

Volume = 43,120 gal
Dimensions: 21' x 25' x 12'

Filter Pumps

Feed Pumps (2): vertical, radial flow, 400 gpm
Backwash Pumps (2): vertical, radial flow, 800 gpm

Filters

Type (2): Downflow, pressure
Size: 7' diameter, 36" bed depth

Carbon Adsorber Feed Sump

Volume = 35,640 gal
Dimensions: 18' x 25' x 12'

Carbon Adsorber Pumps

Feed Pumps (2): vertical, radial flow, 400 gpm
Backwash Pumps (2): vertical, radial flow, 1200 gpm

Carbon Adsorbers

Adsorbers (3): 10' diameter, 20' high
Adsorber Carbon Inventory = 53,694 lb
Carbon Exhaustion Rate = 725 lb/day
Virgin Carbon Storage = 640 ft^3 (19,200 lb)
Spent Carbon Storage = 640 ft^3
Regenerated Carbon Storage: in third adsorber

Adsorber Backwash and Effluent Sump

Volume = 47,120 gal
Dimensions: 23' x 25' x 12'

A carbon adsorption system may possibly be leased on a contract basis.

Flash Mix Chamber (Disinfection)

Volume = 285 gal
Dimensions: 3.5' x 3.5' x 4'
Mixer: 2 BHP
Chlorine Dosage = 1-5 mg/l Cl_2
= 3.4-17 lb Cl_2/day

Distribution Tank

Volume = 297,000 gal
Renovate old aerobic digester.

Reclaimed Wastewater Storage Basin

Volume = 1.5 $\bar{\text{M}}$G

Off-Specification Basin

Volume = 1.5 $\bar{\text{M}}$G

These design criteria are derived in Chapter VII of this report.

3. Develop a contingency plan for removal of inorganic dissolved solids during initial operations of the recirculation scheme.

CHAPTER II

INTRODUCTION

The Owens-Corning Fiberglas Corporation (OCF) in 1968 established an objective of total water recycle for all its manufacturing facilities. The technology of total recycle was successfully developed and demonstrated for fibrous glass insulation facilities in 1968 at the Corporations's Barrington, New Jersey plant. Following the success in that area, the Corporation set out to develop the technology of total recycle for textile fibrous glass facilities. After conducting the necessary background research, OCF submitted an application for federal funding.

The Corporation was awarded an Environmental Protection Agency Demonstration Grant (S801173) in March, 1973. Research and development work (Phase I) for the grant was conducted at the Owens-Corning Fiberglas Corporation manufacturing facility at Anderson, South Carolina. This work along with construction work under Phase II of the grant were originally to be completed over a three-year period. Research and development work delays, brought about by interruptions in manufacturing operations at the Anderson facility, required that the grant period be extended through December 31, 1977.

A detailed work plan for Phase I research and development was defined during the initial months of the grant period. The Corporation at that time retained the firm of Engineering-Science, Inc. to perform the required field studies and preliminary engineering work.

The research and development phase was divided into a series of distinct tasks, which are listed in Table II-1.

TABLE II-1

RESEARCH AND DEVELOPMENT PHASE TASK OUTLINE

Task A	Project Planning
Task B	Project Administration and Reporting
Task C	Establish Current Plant Water Balances, Water Quality, and Material Balances
Task D	Investigate Water Reduction and/or Recycle Alternatives
Task E	Evaluate Existing Wastewater Treatment Systems
Task F	Establish Water Reuse Criteria and Demand
Task G	Evaluate Candidate Water Renovation Processes
Task H	Evaluate Sludge Handling Systems
Task I	Establish Design Criteria and Economics
Task J	Establish Optimum Water Recirculation System

Phase I field studies began in August, 1973, and continued through January, 1974 when grant studies were temporarily suspended due to reconstruction of plant manufacturing facilities. The project field work, including confirmatory survey work, treatability studies, and pilot process recirculation cooling trials, recommenced in June, 1974 and continued through December, 1974 when grant studies were again temporarily suspended due to a significant decrease in plant production rates.

Decreased production rates continued through July, 1975. During this period primary chemical addition systems and Chemical Factory wastewater surge handling facilities were added to the existing wastewater treatment facility, both of which greatly improved removal efficiencies and stabilized performance. Beginning in July, 1975 and continuing through May, 1976, the pilot process recirculation cooling trials were completed.

Research and development data were evaluated during May, 1976. Total recirculation of industrial wastewater at the Anderson facility was determined to be technically and economically feasible.

Following this decision, work began towards design of the recirculation system. A schedule for completion of this work and operation of the facility is as follows:

July 1, 1976	Submit preliminary engineering report to South Carolina and EPA Industrial Environmental Research Lab
July 1, 1977	Advanced wastewater treatment facilities in operation
December 31, 1977	Wastewater recirculation system in operation
January 1, 1978- September 30, 1978	Evaluation of full-scale total recycle system

CHAPTER III

PLANT WATER AND WASTEWATER CHARACTERIZATION

Establishment of an overall water and wastewater balance for the Anderson Plant is of paramount importance in devising plans for water recirculation and designs for advanced wastewater treatment processes. This chapter begins with descriptions of the wastewater surveys that have been conducted at the plant and concludes with projected flows for the recirculation system. A revised sewer plan for the Anderson Plant was established and sample points for the 1973-1974 wastewater survey and characterization study were selected. These will be discussed later in the chapter.

WASTEWATER SURVEY - 1969

During a two-week period in August and September, 1969, T.V. Powers, Jr., a project engineer at the Anderson Plant, conducted a survey of the industrial and sanitary wastewaters in the plant. Flow measurements were taken with triangular weirs and by timed volume; flow readings were taken throughout each day of the survey at two hour intervals. Even though no wastewater characterization analyses were performed, each source was identified with respect to location and visual characteristics of the wastewater. Results from the survey are presented in Table III-1. The closure between the sum of the measured individual flow (310 gpm) and the flow through the treatment plant (466 gpm) is relatively poor, amounting to a difference of 156 gpm. During the survey approximately 84 gpm of wastewater were being bypassed around the treatment plant.

WASTEWATER SURVEY AND MASS BALANCE - 1973 and 1974

A comprehensive water and wastewater survey and characterization study was conducted by ES during the fall of 1973 and again in October, 1974 (following Factory "D" rebuild). Each process and sanitary wastewater source was identified prior to the survey and labeled, as described in Table III-2. Additionally, the waters in the various cooling systems were characterized in order to assess requirements for the use of reclaimed water. Several methods of flow measurement

TABLE III-1

WASTEWATER SURVEY - 1969

DESCRIPTION	1969 SAMPLE NO.	1973-1974 SAMPLE NO.	MEAN FLOW, gpm	
"A" Factory				
Sanitary Wastewater	6	03	20	
Air Conditioning	7	01	5	
Binder Washdown	8		3	
Forming Washdown and Air Wash	11		34	
Air Wash	14		2	
Ceramic Saw	9	None	0	
Sanitary Wastewater	15	02	5	
TOTAL "A" FACTORY = A			69	
Beta Factory				
Sanitary Wastewater	4	12	4	
Air Wash	5	10	15	
Forming Washdown	10	11	35	
Binder Washdown	12		18	
TOTAL BETA FACTORY = B			72	
TOTAL GRAVITY LINE TO TREATMENT PLANT = C = A + B				141
"D" Factory				
Forming Wash	24	30	47	
Binder Wash	25		16	
TOTAL "D" FACTORY = D			63	
Chemical Factory				
All	17	50	39	
TOTAL CHEMICAL FACTORY = E			39	
TOTAL "D" FACTORY FORCE MAIN = F = D + E				102
Mat Line				
All	3	61	67	
TOTAL MAT LINE = G			67	
TOTAL WASTEWATER TO TREATMENT PLANT = C + F + G				310
TOTAL WASTEWATER THROUGH TREATMENT PLANT (as measured)				466

TABLE III-1 (Continued)

WASTEWATER SURVEY - 1969

DESCRIPTION	1969 SAMPLE NO.	1973-1974 SAMPLE NO.	MEAN FLOW, gpm	
Bypass				
Alloy	1	06	19	
No. 1 Spray Pond	2	62	0	
Surface	16	22	4	
Surface - Chemical Factory	22	43	45	
"D" Factory Surface	23	36,37,38	16	
TOTAL WASTEWATER BYPASSING TREATMENT			84	
TOTAL WATER PURCHASED				766

TABLE III-2

1973-1974 WASTEWATER SURVEY SAMPLE POINT DESCRIPTION

SAMPLE POINT NUMBER(s)	DESCRIPTION	TYPE OF FLOW MEASUREMENT	WATER QUALITY ANALYSIS
	"A" Factory and Office Building		
01	Combined Wastewater Flow from Basement Floor Trench of "A" Factory (Process)	4" Cippoletti Weir w/ Stevens Recorder	yes
02	Waste Flow from First Floor East Side of "A" Factory (Sanitary)	Timed Volume	yes
03	Waste Flow from "A" Factory Kitchen, etc. West Side (Sanitary)	Timed Volume	yes
05, 82	AC System Blowdown "A" Factory Basement (Chill Water to #1 Spray Pond)	Water Meter	no
06	Storm Water Collection Outfall West of "A" Factory (includes Alloy Wastes)	90° "V" Notch Weir w/ Stevens Recorder	yes
	Beta Factory		
10	Beta Factory Air Scrubber Blowdown (Basement)	90° "V" Notch Weir No Recorder	yes
11	Beta Factory Basement Floor Trench Less No. 10 (Process)	90° "V" Notch Weir w/ Stevens Recorder	yes
12	Sanitary Wastewater Sewer North of Beta Building	Estimate	no
13	Makeup Chilled Water System Beta Factory	Timed Volume	no
	Spray Pond No. 1 and 2		
20	No. 1 Spray Pond Water (For Blowdown See No. 62)	None	yes

TABLE III-2 (Continued)

1973-1974 WASTEWATER SURVEY SAMPLE POINT DESCRIPTION

SAMPLE POINT NUMBER(s)	DESCRIPTION	TYPE OF FLOW MEASUREMENT	WATER QUALITY ANALYSIS
21	No. 2 Spray Pond Water	None	yes
22	No. 2 Spray Pond Blowdown	See No. 24	yes
23	No. 2 Spray Pond Filter Backwash	See No. 24	yes
24	Combination of No. 22 and No. 23	90° "V" Notch Weir w/ Stevens Recorder	no
	"D" Factory and Tire Cord Building		
30	Combined Wastewater Flow from "D" Factory, Includes Process and Scrubber Blowdown	45° "V" Notch Weir w/ Stevens Recorder	yes
31	"D" Factory Air Scrubber Blowdown (Basement)	Estimate	yes
32	"D" Factory Condenser Cooling Tower Water	None	yes
33, 84	"D" Factory Condenser Cooling Tower Blowdown	Water Meter Water Meter	yes
34	"D" Factory Process Cooling Tower Water	None	yes
35, 85	"D" Factory Process Cooling Tower Blowdown	Water Meter	yes
36	"D" Factory Process Cooling Tower Filter Backwash (12" Clay Pipe)	Estimate	no
37	24" RCP Storm Water North of "D" Factory Cooling Towers	Timed Volume	no
38	18" Steel Pipe North of "D" Factory Cooling Towers	Timed Volume	no

TABLE III-2 (Continued)

1973-1974 WASTEWATER SURVEY SAMPLE POINT DESCRIPTION

SAMPLE POINT NUMBER(s)	DESCRIPTION	TYPE OF FLOW MEASUREMENT	WATER QUALITY ANALYSIS
39	Surface Outfall of Caustic Wash Building East of "D" Factory	Estimate	no
40	Sanitary Waste from "D" Factory Whse.	Estimate	no
41	Sanitary Waste from Tire Cord Lift Station	6" Cippoletti Weir w/ Stevens Recorder	no
42	Latex Pit Overflow	Timed Volume	no
43	Surface Runoff Basin Sluice Gate East of Tire Cord Area	Weir	no
	Chemical Factory		
50	Total Waste from Chemical Factory	Timed Volume	no
51	Chemical Factory Process Cooling Tower Water	None	no
52	Chemical Factory Process Cooling Tower Blowdown	Timed Volume	no
53	Inert Gas Cooling Tower Water	None	no
54	Inert Gas Cooling Tower Blowdown	Timed Volume	no
55	Surface Runoff East of Plant	Weir	no
56	Inert Gas Scrubber Effluent	Timed Volume	yes
	Wastewater Treatment Plant		
60	Treatment Plant Influent Exclusive of Mat Line	12" Cippoletti Weir w/ Stevens Recorder	yes

TABLE III-2 (Continued)

1973-1974 WASTEWATER SURVEY SAMPLE POINT DESCRIPTION

SAMPLE POINT NUMBER(s)	DESCRIPTION	TYPE OF FLOW MEASUREMENT	WATER QUALITY ANALYSIS
61	Mat Line to Treatment Plant	90° "V" Notch Weir w/ Stevens Recorder	yes
62	Sewer West of Treatment Plant Including No. 1 Spray Pond Blowdown and Wastes from Garage and Caustic Wash Facility	90° "V" Notch Weir w/ Stevens Recorder	yes
63	Waste Treatment Plant Effluent	Parshall Flume	yes
64	Waste Treatment Plant Influent	None	yes
	Cooling Systems		
80	No. 1 Spray Pond Makeup Water	Water Meter	no
81	"A" Factory Chill Water Makeup	Water Meter	no
83	No. 2 Spray Pond Makeup Water	Water Meter	no
86	"D" Factory Chill Water Makeup	Water Meter	no
87	"D" Factory Chill Water Blowdown to "D" Factory Condenser Water Cooling Tower	Water Meter	no
88	"D" Factory Condenser Water Cooling Tower Makeup	Water Meter	no
89	"D" Factory Process Water Cooling Tower Makeup	Water Meter	no
90	"D" Factory Air Scrubber Makeup	Water Meter	no
91	Chemical Factory Cooling Tower Makeup - Cell #1 (East)	Water Meter	no

TABLE III-2 - (Continued)

1973-1974 WASTEWATER SURVEY SAMPLE POINT DESCRIPTION

SAMPLE POINT NUMBER(s)	DESCRIPTION	TYPE OF FLOW MEASUREMENT	WATER QUALITY ANALYSIS
92	Chemical Factory Cooling Tower Makeup - Cell #2 (West)	Water Meter	no
93	Chemical Factory Inert Gas Scrubber Cooling Tower Makeup	Water Meter	no
94	City Water Meter Supplying OCF Plant	Water Meter	yes
95	Chemical Factory Carrier System Makeup	Water Meter	no
96	Boiler Water Makeup	Water Meter	no

were utilized; as listed in Table III-2, these included weirs with recorders, water meters and timed volumes.

Wastewater Flows and Mass Balance

Flow and mass balances for the process and sanitary wastewaters are listed in Table III-3. Comparison of the sum of the tributary flows to the measured treatment plant influent flow indicates that the flow closure is excellent. The major contributors to wastewater flow are "A" Factory, Chemical Factory, and the Mat Line. Approximately 44 gpm of the measured flow was sanitary wastewater from the plant restrooms and cafeteria. It should be noted that the tributaries which were bypassing the wastewater treatment facility during the 1969 survey had been routed to treatment prior to the 1973-1974 survey. Therefore, all flows except some uncontaminated storm water runoff now receive treatment prior to discharge.

Major contributors to wastewater COD, TSS, and TDS were the fibrous glass manufacturing process sources in "A" Factory, Beta Factory, and "D" Factory. Contamination is a result primarily of spills and leakages of binder used in the fiberglass manufacturing processes. The discrepancy between the COD mass in the summation of the tributaries and that in the treatment plant influent is a result of aperiodic binder spills not represented in the composited grab samples at each source but included in the equalized treatment plant influent. While the mass closures for TSS and TDS were not as good as those for COD, the total solids closure (total solids = TSS + TDS) is reasonable, since the various binders used at Anderson tend to precipitate from solution when mixed.

Heating and Cooling Systems

Water flows and characterization analyses for the heating and cooling systems at the Anderson Plant were determined during the 1973-1974 survey in order to enumerate water reuse requirements. Makeup, blowdown, drift, and evaporation flows for both fall and winter operations were measured and/or calculated through the use of water meters on makeup piping and water quality analyses (TDS, SiO_2, Total Hardness) on waters in the systems. While the results are not presented here, since future conditions will be much different than those in 1973 and 1974 as a result of planned

TABLE III-3

1973-1974 WASTEWATER SURVEY - MASS BALANCE

PROCESS AREA	ES SAMPLE NO.	MEAN FLOW (gpm)	PERCENT OF TOTAL	COD (mg/l)	COD (lb/d)	PERCENT OF TOTAL	TDS (mg/l)	TDS (lb/d)	PERCENT OF TOTAL	TSS (mg/l)	TSS (lb/d)	PERCENT OF TOTAL
"A" Factory Process	01	60	20	2565	1848	32.5	1079	778	31	568	409	34.4
"A" Factory Sanitary East End	02	6	2	630	45	0.8	842	61	2	184	13	1
"A" Factory Sanitary West End	03	28	9	359	121	2.1	192	65	3	58	20	2
Beta Factory Scrubber	10*	12	4	-	-	-	-	-	-	-	-	-
Beta Factory Process	11	15	5	3443	620	10.9	1806	325	13	556	100	8
Beta Factory Sanitary	12*	4	1	-	-	-	-	-	-	-	-	-
Chemical Factory Process	50	33	11	684	271	4.8	189	75	3	197	78	7
Chemical Factory Inert Gas Scrubber	56	26	8	15	5	0.1	66	21	1	6	2	0.2
"D" Factory Scrubber	31*	2	1	-	-	-	-	-	-	-	-	-
"D" Factory Process	30	23	8	8279	2287	40.2	1596	441	17	1093	302	25
"D" Factory and Tire Cord Sanitary	40, 41	6	2	290	21	0.4	192	14	0	61	4	0.3
Alloy Process	06	31	10	65	24	0.4	189	70	3	51	19	2
Power House	62	6	2	957	69	1.2	6044	436	17	294	21	2
No. 2 Spray Pond Filter Backwash	24	4	1	37	2	0	279	13	0	43	2	0.2
Miscellaneous	37,38 & 39	10	3	62	7	0.1	131	16	1	68	8	0.7
"D" Factory Filter Backwash	36*	4	1	-	-	-	-	-	-	-	-	-
Estimated Content of Unsampled Flows	*	-	-	71	19	0.3	129	34	1	66	17	1
Subtotal		270	88	-	5339	94	-	2349	92	-	995	84
Mat Line	61	35	12	834	351	6	470	198	8	463	195	16
Total Tributaries		305	100	-	5690	100	-	2547	100	-	1190	100
Measured Treatment Plant Influent	63,64	303	-	2001	7281	-	336	1223	-	572	2081	-

* Waste streams not sampled for water quality

manufacturing changes in "A" Factory and Beta Factory and the addition of "E" Factory, they do form the basis for the summer and winter cooling system flow projections given later in this chapter.

WASTEWATER SURVEY - 1976

Confirmatory survey work was conducted by OCF environmental personnel during a one week period in April, 1976. This survey served two purposes; first to determine if changes in wastewater flow rates had occurred since the survey in 1973 and 1974, and second, to aid in the projection of future wastewater flows. Flow rates were measured at 17 points in the sewer system through the use of weirs with recorders, timed volumes, tracer-dilution methods, and a Parshall flume.

Survey results are presented in Table III-4. The flow closure is quite reasonable for a survey of this nature. The measured treatment plant influent flow rate indicates a 38 gpm reduction from that recorded during the 1973-1974 survey. This reduction is due to reduced levels of manufacturing operations in "A" Factory and Beta Factory during the 1976 survey. The flow measurements are not, however, representative of future flows because of the additional changes to occur in "A" Factory and the additional manufacturing area which is to be constructed ("E" Factory).

PROJECTED COOLING SYSTEM WATER BALANCE

Water balances for the several cooling systems in the plant are a prerequisite for establishing water reuse patterns. Projected cooling water flows for summer and winter operations are presented in this section. These projections are based upon:

- Measured and calculated cooling flows for fall and winter operations as determined during the survey in 1973-1974.
- Planned heat load reductions in "A" Factory and Beta Factory; future heat load additions from "E" Factory.
- Plant operating records.

TABLE III-4

WASTEWATER SURVEY - 1976

PROCESS AREA	1973-1974 SAMPLE NO.	MEAN FLOW (gpm)
"A" Factory Process	01	25
"A" Factory Sanitary	02, 03	16
Beta Factory Process, Binder, & Sanitary	10, 11, 12	15
Chemical Factory Process	50	29
"D" Factory Process & Scrubber	30, 31	29
Chemical Factory I.G. Scrubber	56	41
"D" Factory Sanitary	40, 41	6
Alloy	06	16
Powerhouse	62	2
Filter Backwash	24, 36	1
Mat Line	61	29
Miscellaneous	37, 38, 39	46
TOTAL		255 gpm
Measured Flow Through Treatment Plant	63, 64	265 gpm

Reuse Scheme

Basically, the plan for wastewater recirculation/reuse in the cooling systems is as follows:

- Upgrade existing wastewater treatment facilities to enable production of an effluent of such quality as may be used in the plant cooling systems.
- Utilize this reclaimed water as makeup to the cooling systems.
- Cascade blowdowns from one cooling system to another; thus, in effect, the blowdowns will be part of the makeup to the systems receiving them.
- Final blowdowns from the cooling systems are to be routed to the "D" and "E" scrubbers.

Water Balance

There are nine major cooling systems in the Anderson Plant which use water; these are described in Table III-5. The process cooling systems require the highest degree of water quality due to extreme heat loads and small diameter distribution piping. The chillers possess somewhat liberal physical and chemical water quality requirements, but the water used must be free of pathogenic bacteria and viruses, because the chill water is used to cool the atmosphere inside each factory.

Cooling of heated circulating water in evaporative cooling systems employed at the Anderson Plant is accomplished primarily through evaporation in the spray ponds and cooling towers. Some water is also lost through entrainment of water droplets in air draft; this loss is known as "drift". While both evaporation and drift constitute water vapor losses, of the two mechanisms only the drift process is responsible for dissolved solids removal. The overall effect is that dissolved solids concentrate in the remaining liquid. To prevent a buildup of dissolved solids (and associated scaling and heat transfer problems) in the cooling system, a small portion of the circulating water is continuously discharged to the treatment system; this loss is termed "blowdown". The end result is that water is continuously added to each system (makeup) in amounts equal to the total water lost (blowdown + evaporation + drift).

TABLE III-5

MAJOR COOLING SYSTEMS

SYSTEM	PURPOSE	VOLUME (gal)
"A" Chillers	Cools water for "A" Factory and Beta Factory air washers	20,000
"E" Chillers (Future)	Cools water for "E" Factory air washers	38,000*
"D" Chillers	Cools water for "D" Factory air washers	38,000
#1 Pond ("A" & "E" Condenser Cooling)	Cools refrigeration units in "A" Factory, Beta Factory, and "E" Factory	755,000
#2 Pond ("A" & "E" Process Cooling)	Cools bushings, fin shields, and furnace coils in "A" Factory, Beta Factory, and "E" Factory	505,000
"D" Condenser Cooling	Cools refrigeration units in "D" Factory	38,000
"D" Process Cooling	Cools bushings, fin shields, and furnace coils in "D" Factory	135,000
Chemical Cooling Tower No. 2	Cools Chemical Factory process units: #2 Thinning Tank, #2, #3, and #4 Reactors, and Carrier chill water condensers	6,000
Chemical Cooling Tower No. 1	Cools burner in the Chemical Factory Inert Gas manufacturing operation, #1 Reactor, and Trane chill water condensers	2,500
	TOTAL	1,537,500

* Estimated volume

Makeup, blowdown, evaporation, and drift flows for both summer and winter operating conditions for all nine major cooling systems are illustrated in Figures III-1 and III-2. These figures also depict the reclaimed water requirements and cascade pattern. The totals for operations during both seasons are listed in Table III-6.

Drift Losses

Mass balances have been performed on each of the cooling systems in accordance with the flows shown in Figures III-1 and III-2. These balances may serve several functions:

- calculation of equilibrium levels of any reclaimed water constituents in each of the cooling systems.
- quantification of drift losses for any constituent.
- determination of requirements for treatment of blowdown.

The first step in this evaluation was the determination of the number of cycles of concentration (C) that will occur in each cooling system. The "C" value is a dimensionless number which expresses the number of times the concentration of any constituent is multiplied from its original value in the makeup water, and is calculated as follows:

$$C = \frac{B + D + E}{B + D}$$

where:

C = cycles of concentration
B = blowdown rate
D = drift loss rate
E = evaporation rate

For example, if the TDS concentration in the makeup water to a particular cooling system is 50 mg/l, and the "C" value for that system is 5.00, the TDS concentration in the system (and in the blowdown and drift) will be 5.00 x 50 mg/l, or 250 mg/l. Cycles of concentration for each of the major cooling systems are listed in Table III-7; these values are based on the projected flows shown in Figures III-1 and III-2.

TABLE III-6

COOLING SYSTEMS SUMMARY

	SUMMER (gpm)	WINTER (gpm)
Reclaimed Water Makeup	174.3	89.6
Blowdown to Treatment	1	1
Blowdown to Scrubbers (to Treatment)	27	3
Drift	14.8	8.8
Evaporation	131.5	76.8

TABLE III-7

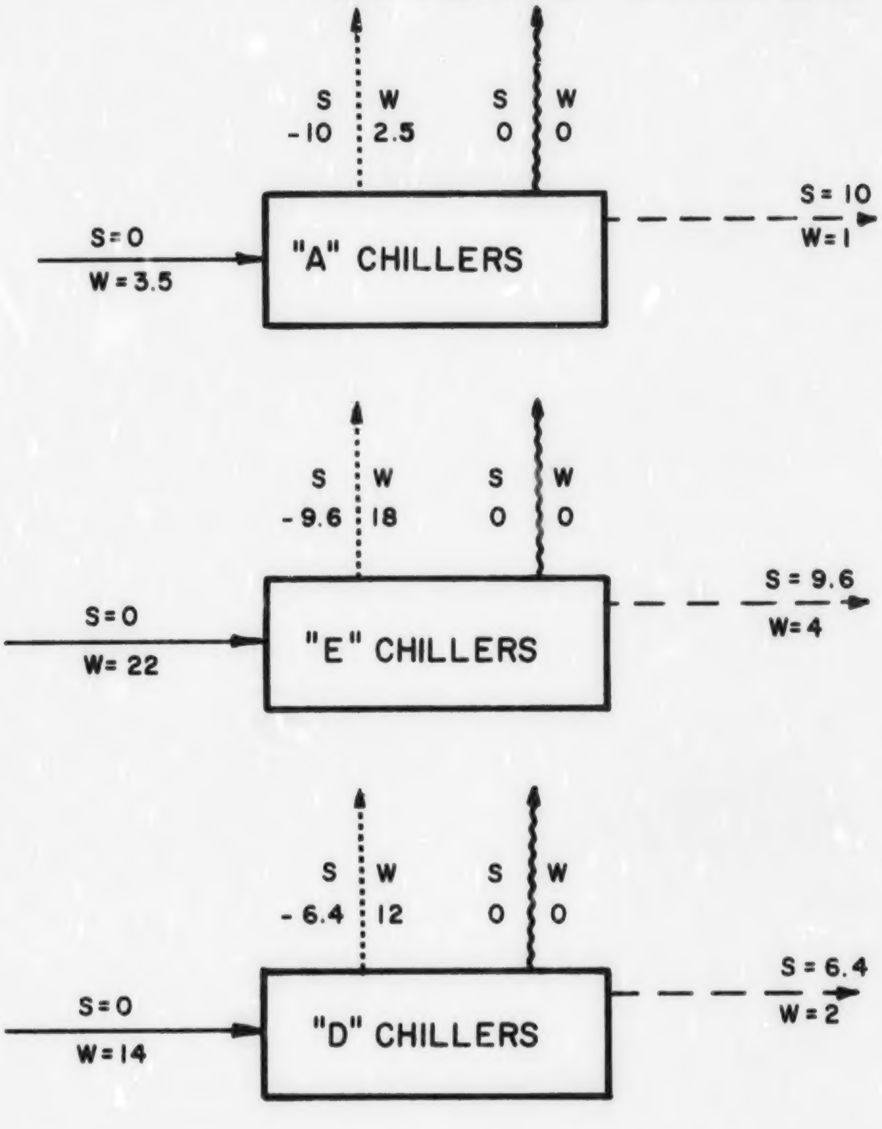
PROJECTED

CYCLES OF CONCENTRATIONS FOR THE COOLING SYSTEMS

SYSTEM	CYCLES OF CONCENTRATION	
	SUMMER	WINTER
"A" Chillers	1.00	3.50
"E" Chillers	1.00	5.50
"D" Chillers	1.00	7.00
#1 Pond ("A" & "E" Condenser Cooling)	4.71	2.00
#2 Pond ("A" & "E" Process Cooling)	4.70	5.25
"D" Condenser Cooling	5.00	1.00
"D" Process Cooling	4.57	8.42
Chemical Cooling Tower No. 2	5.83	4.83
Chemical Cooling Tower No. 1	2.33	1.83

FIGURE III-1

COOLING SYSTEM



Note: Negative Evaporation Values Indicate Condensation Of Humidity In Air

FIGURE III-1

WATER BALANCE

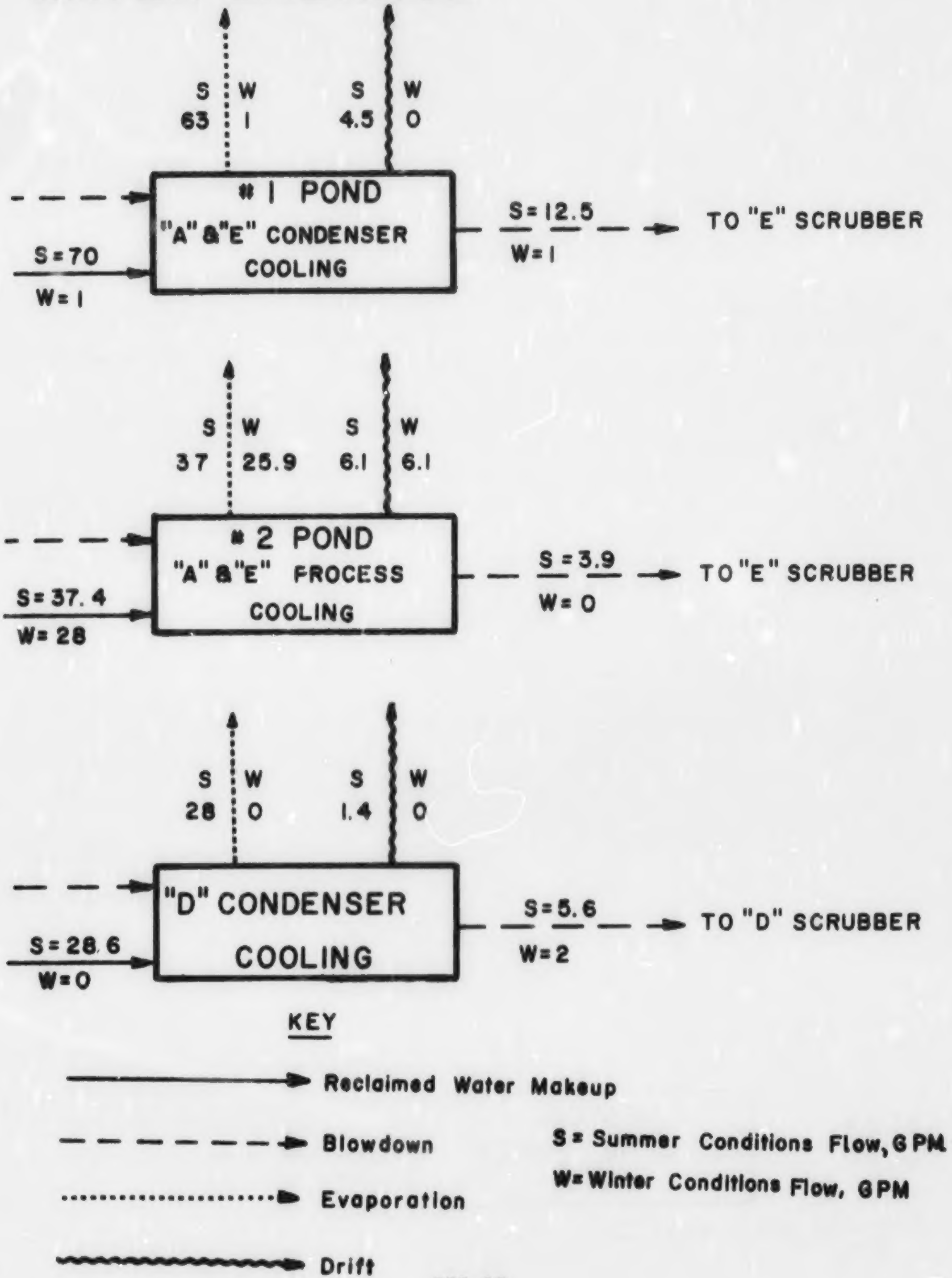
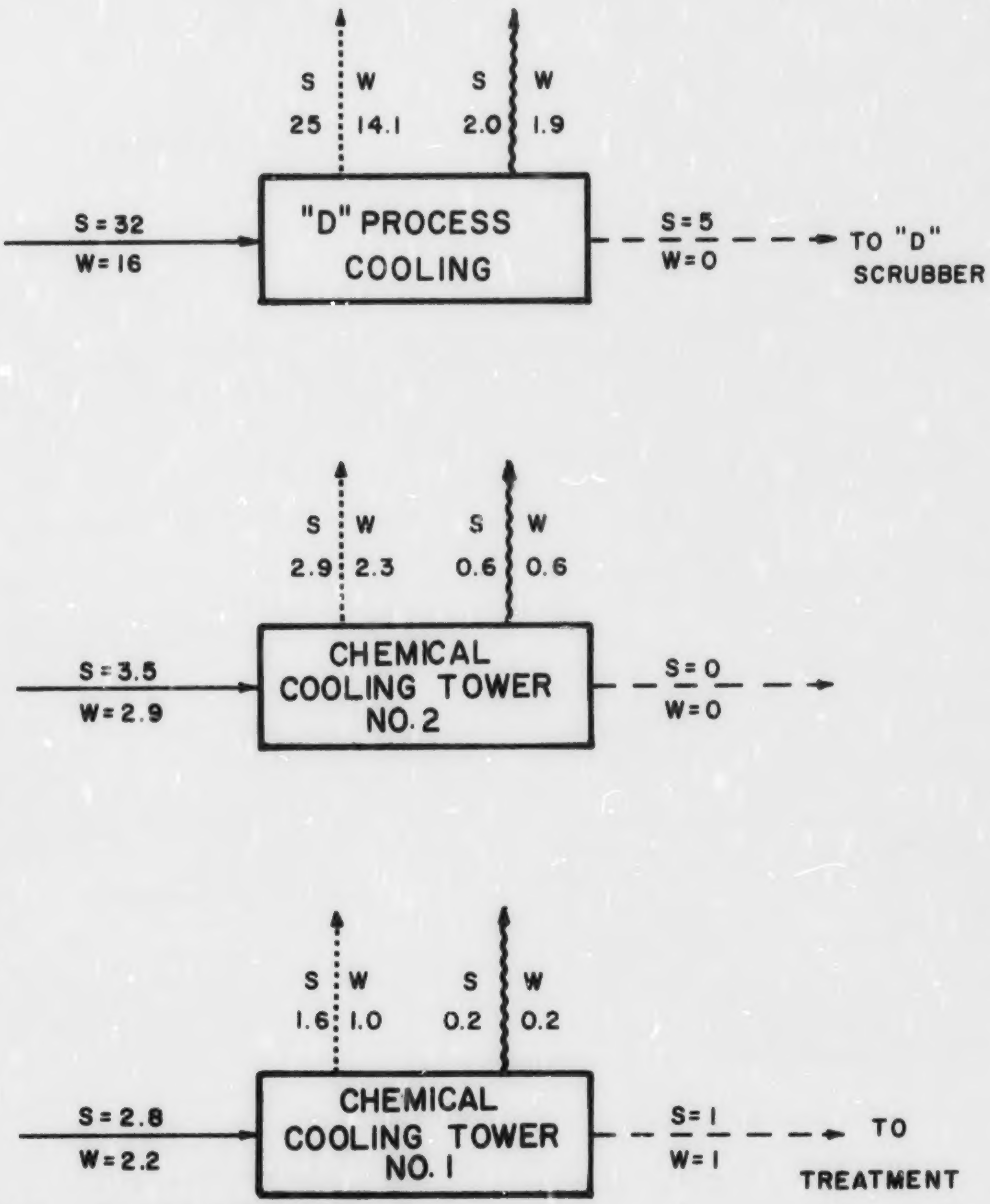


FIGURE III-2

COOLING SYSTEM WATER BALANCE



NOTE: FOR KEY SEE FIGURE III-1

The next step was to calculate mass balances around each of the cooling systems in terms of the concentration (W) of any constituent in the reclaimed makeup water, incorporating the "C" values calculated earlier. Drift losses are summarized in Table III-8. An application of these results is presented in the last section of this chapter.

PROJECTED WASTEWATER FLOWS AND REUSE

Projected wastewater flows and reuse requirements have been developed from cooling system balances given earlier in the chapter, process reuse flows developed in 1974, planned manufacturing reductions in "A" Factory and Beta Factory, and future manufacturing additions ("E" Factory). Sanitary wastewaters from the entire Anderson Plant, amounting to 40-60 gpm, will be segregated from the process wastewater conveyance system, treated separately in a "package" plant, and discharged to Betsy Creek. These wastewaters originate from potable city water uses in the plant. Usages requiring high quality water, such as boiler makeup, binder makeup, and deionized water sprays will not utilize reclaimed wastewater. Finally, reclaimed wastewater will be used in the cooling systems, the air scrubbers, Mat Line, Alloy, and for floor washdowns in "A", "D", and "E" Factories. Although current plans rely upon using city water as input to the IG Scrubber and then using Scrubber effluent in the Mat Line ("piggyback" system), the possibility of developing an internal recycle system (with city water) for the IG Scrubber is being investigated. In that case, reclaimed water would be used for the Mat Line. The most sensitive of these uses, which is also the largest use, is the cooling systems. Accordingly, a pilot cooling loop has been tested using reclaimed wastewater, as will be discussed in Chapter VI.

Both cooling and process reuse flows, together with wastewater discharges, are listed in Table III-9. Since cooling system evaporative losses will be much greater during the summer than during the winter, summer reuse requirements will exceed the amount of reclaimed wastewater available; conversely, during the winter, reclaimed wastewater flows will exceed reuse requirements. The obvious conservation solution would be to store the excess reclaimed water during the winter for later

TABLE III-8

PROJECTED

DRIFT LOSSES FOR THE COOLING SYSTEMS

SYSTEM	DRIFT LOSSES* SUMMER (lb/day)	WINTER (lb/day)
"A" Chillers	------	------
"E" Chillers	------	------
"D" Chillers	------	------
#1 Pond ("A" & "E" Condenser Cooling)	0.254W	0
#2 Pond ("A" & "E" Process Cooling)	0.344W	0.600W
"D" Condenser Cooling	0.084W	0
"D" Process Cooling	0.110W	0.192W
Chemical Cooling Tower No. 2	0.042W	0.035W
Chemical Cooling Tower No. 1	0.006W	0.004W
TOTALS	0.840W	0.831W

---

* Expressed in terms of W, the concentration of the constituent in the reclaimed water.

TABLE III-9

PROJECTED WASTEWATER FLOWS AND REUSE FOR RECIRCULATION

SOURCE	FLOW (gpm)	SUMMER REUSE (gpm)	WINTER REUSE (gpm)
"A" Factory Process	15	10	10
Marble Factory/Binder Room	2	0	0
Chemical Factory Process	35	0	0
"D" Factory Process	25	20	20
"D" Scrubbers[1]	11	0.4	9
"E" Factory Process	25	20	20
"E" Scrubbers[1]	20	3.6	19
Alloy	20	20	20
IG Scrubber - Mat Line[2]	26	0	0
Boilerhouse	5	0	0
Filter Backwash (3)	1	1	1
Miscellaneous	19	10	10
Cooling Systems[3]	1	174.3	89.6
TOTALS	205	259.3	198.6

NOTE:
1. Both scrubbers receive cooling system blowdown in addition to flows listed in reuse columns.
2. "Piggyback" system.
3. Wastewater flow is Chemical Cooling Tower No. 1 system blowdown; reuse flows are totals for all cooling systems.

use in the summer. However, at this time it is not possible to determine the balance between those days with excess wastewater and those with deficit. The "safest" path, therefore, is to make provisions for controlled addition of city water makeup to the reclaimed wastewater distribution tank, and to ensure that the reclaimed water storage basin is of sufficient volume to hold excess flows for at least 100 days (the entire winter season). An overall recirculation water balance for the Anderson Plant is shown as Figure III-3.

PROJECTED EQUILIBRIUM CONCENTRATIONS FOR INORGANIC CONTAMINANTS

Drift loss values presented in Table III-8 along with recent water and wastewater characterization data have been used to project equilibrium concentrations of hardness (total and calcium), silica, sulfate, and zinc in the reclaimed wastewater. These parameters were chosen for analysis because their concentrations in the cooling systems have been limited, as will be described in Chapter VI. Projections were calculated utilizing data contained in Table III-10. The methodology used in this analysis is illustrated (for total hardness) in the following paragraphs.

As a first step, calculate the daily input of total hardness to the plant waters (assume negligible cooling input). Process input is due to hardness in the city water and hardness added through process uses. Quality data in Table III-10 show that total hardness increases by 25 mg/l in each pass through the manufacturing facilities; city water contains approximately 16 mg/l total hardness. Based upon the flow rates in Figure III-3, the daily total hardness input is equal to the sum of the city water input [92 gpm @ (16 mg/l + 25 mg/l)] and the reclaimed wastewater input [113 gpm @ 25 mg/l]

City Water Input =

$$\left(92\ \frac{\text{gal}}{\text{min}}\right)\left(\frac{1440\ \text{min}}{\text{day}}\right)\left(\frac{\bar{\text{M}}\text{G}}{10^6\ \text{gal}}\right)(41\ \text{mg/l})\left(\frac{8.34\ \text{lb}/\bar{\text{M}}\text{G}}{\text{mg/l}}\right) = 45.3\ \text{lb/day}$$

Reclaimed Wastewater Input =

$$\left(113\ \frac{\text{gal}}{\text{min}}\right)\left(\frac{1440\ \text{min}}{\text{day}}\right)\left(\frac{\bar{\text{M}}\text{G}}{10^6\ \text{gal}}\right)(25\ \text{mg/l})\left(\frac{8.34\ \text{lb}/\bar{\text{M}}\text{G}}{\text{mg/l}}\right) = 33.9\ \text{lb/day}$$

Total Input = 45.3 lb/day + 33.9 lb/day = 79 lb/day

This input is a net value which includes any removal through treatment processes.

If drift loss is the only mechanism of hardness removal, the concentration of hardness in the reclaimed wastewater will continuously

TABLE III-10

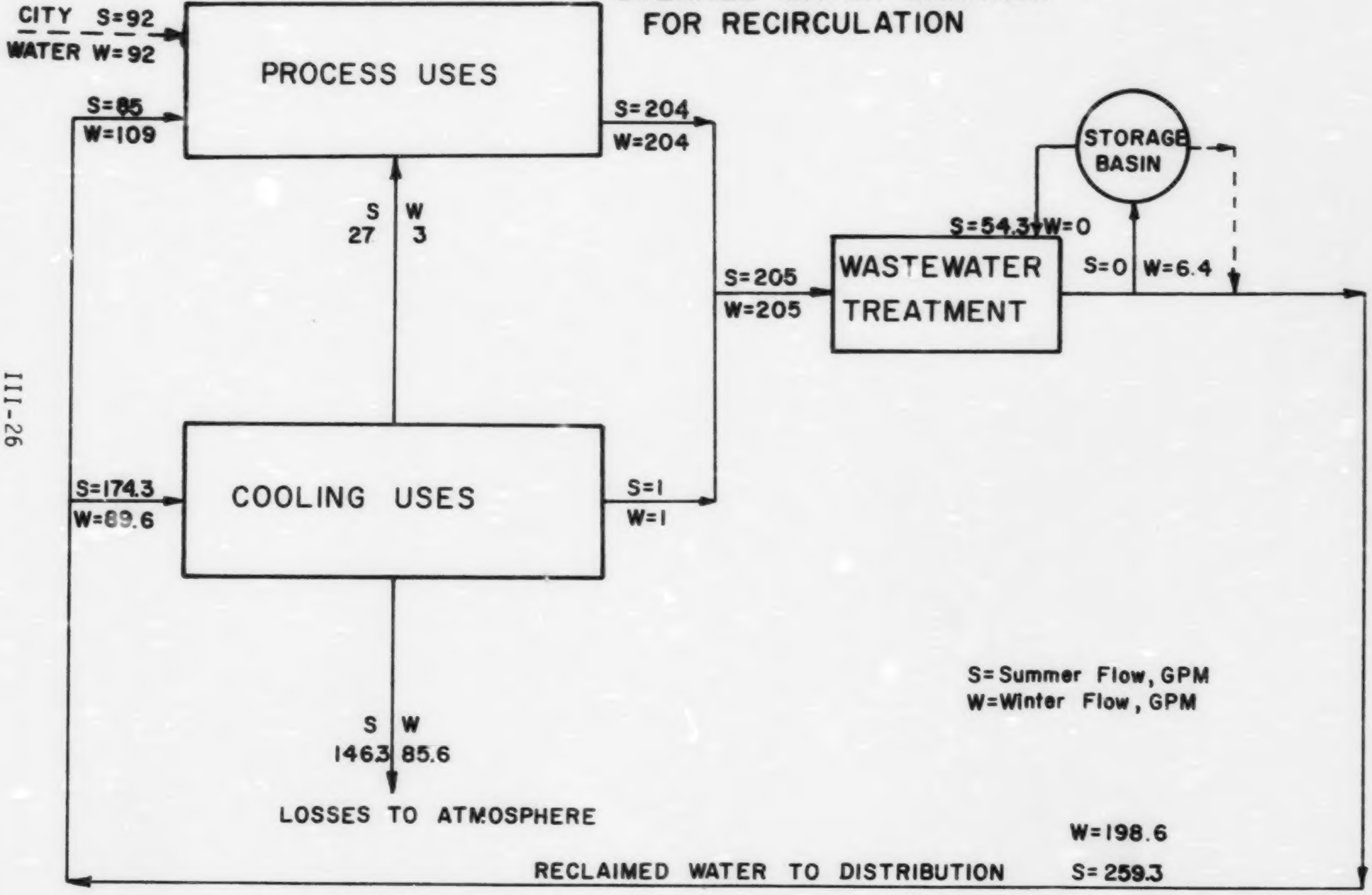
QUALITY OF RECLAIMED WASTEWATER AND CITY WATER

Parameter	Reclaimed Wastewater (mg/1)	City Water (mg/1)	Increase (mg/1)
Total Dissolved Solids	280	52	228
Total Hardness (as $CaCO_3$)	41	16	25
Calcium Hardness (as $CaCO_3$)	35	10	25
Silica (as SiO_2)	19	9	10
Sulfate	74	19	55
Zinc	2.0	0.1	1.9

*Mean values during 90-day pilot cooling loop trial: February, 1976 - April, 1976.

FIGURE III-3

OVERALL WATER BALANCE FOR RECIRCULATION



increase until the mass of hardness lost through drift is equivalent to the net daily input. Using the drift loss totals from Table III-8, the equilibrium hardness levels may be calculated as follows:

Hardness Input = Drift Losses, and for summer conditions

$$79 \text{ lb.day} = 0.840 \text{ W lb/day}$$

$$W = \frac{79}{0.840} \text{ mg/l}$$

$$W = 94 \text{ mg/l}$$

Similarly, for winter conditions

$$79 \text{ lb/day} = 0.831 \text{ W lb/day}$$

$$W = \frac{79}{0.831} \text{ mg/l}$$

$$W = 95 \text{ mg/l}$$

Projected inputs and reclaimed wastewater equilibrium concentrations for total dissolved solids, calcium hardness, silica, sulfate, and zinc were calculated using the format shown above. Results are summarized in Table III-11.

Though the equilibrium concentrations themselves do not appear to be excessive, when the contaminants are concentrated in the cooling systems, the resultant system concentrations may be great enough to cause scaling, plugging, and associated heat transfer problems. Zinc concentrations may be at levels which are biocidal to the activated sludge system. To avert these problems, it will be necessary to remove the majority of the contaminants from the water system. Realizing the potential of the cooling cascade pattern, the logical candidate streams for treatment or discharge are the ones in which dissolved contaminants will be concentrated to the greatest degree. This topic will receive further attention in Chapters VI and VII.

TABLE III-11

PROJECTED INPUTS AND EQUILIBRIUM CONCENTRATIONS FOR INORGANIC CONTAMINANTS

Parameter	Daily Input (lb/day)	Equilibrium Concentrations in Wastewater* Summer (mg/l)	Winter (mg/l)
Total Dissolved Solids	618	736	744
Total Hardness	79	94	95
Calcium Hardness	73	87	88
Silica	35	42	42
Sulfate	156	186	188
Zinc	4.8	6	6

*Assuming drift loss is the sole removal mechanism.

CHAPTER IV

OPTIMIZATION OF EXISTING WASTEWATER TREATMENT FACILITIES OPERATIONS

Existing wastewater treatment facilities are described in the first sections of this chapter, including summaries of EPA evaluations of the facilities. Recent facility improvements are discussed in the next section, followed by the historical record of plant performance and the development of design loadings for advanced wastewater treatment processes.

DESCRIPTION OF EXISTING FACILITIES

Portions of the existing treatment facilities first became operational in 1951; however, the treatment plant was expanded somewhat in 1967. A site plan is shown as Figure IV-1, and the process flow diagram is depicted in Figure IV-2. Following paragraphs provide a narrative description of treatment operations.

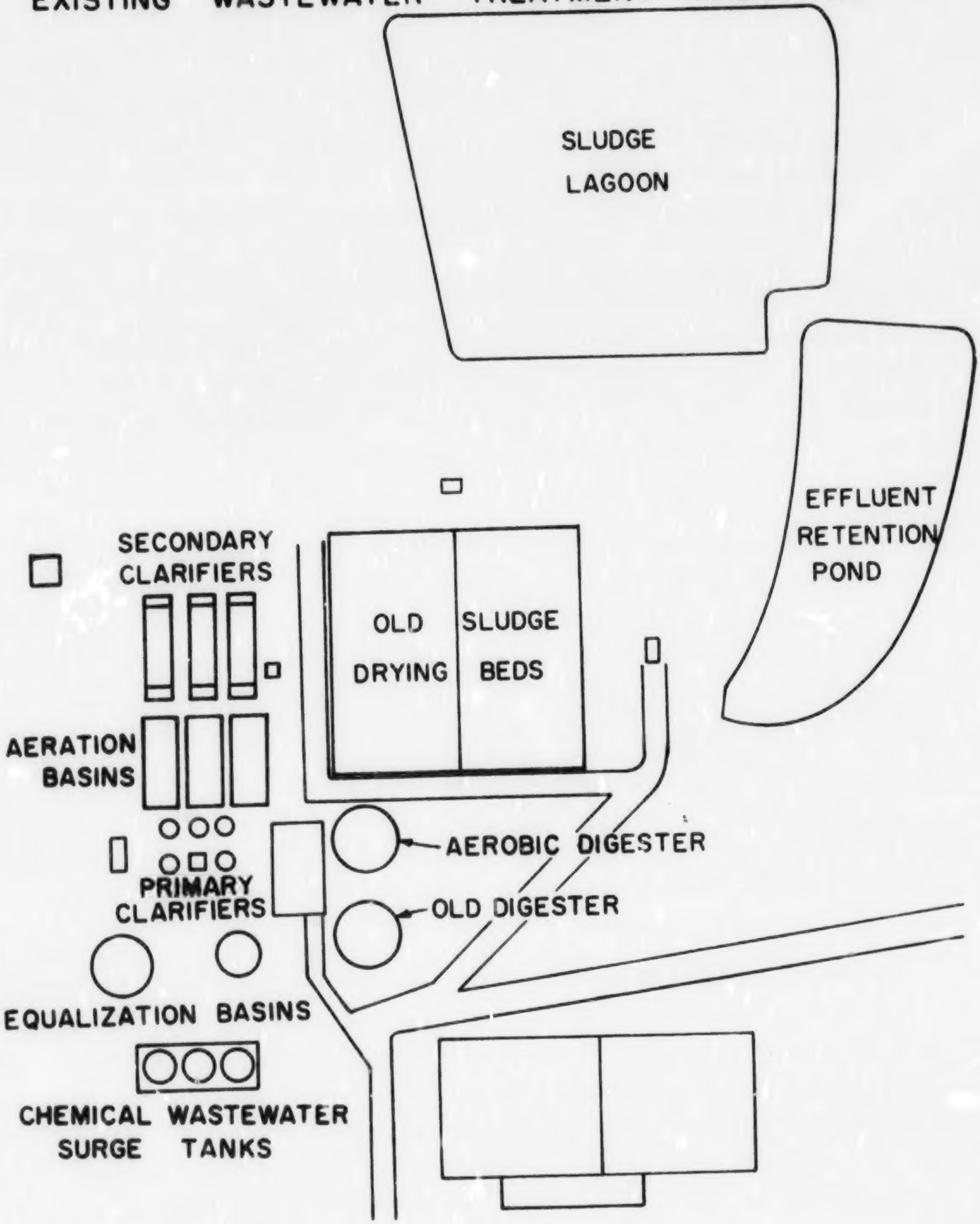
Mat Line wastewater flows through a basket (manually cleaned) which strains some of the fibrous glass strands from the liquid, and then into the small air-agitated equalization basin. Wastewater from "A" Factory, Beta Factory, and "D" Factory flows through a manually cleaned bar screen into the large equalization basin. Chemical Factory wastewater enters the three surge tanks and is bled through the bar screen at a controlled rate. Wastewater is pumped continuously from the large equalization basin (also air-agitated) to the small equalization basin, which overflows into a distribution box. From the distribution box the major portion of the wastewater flows to the flash mix chamber and the remainder flows back into the large equalization basin.

Sulfuric acid, caustic, clay, ferric chloride and polymers are added in the flash mix chamber; flash mix effluent then enters the flocculation tank. A pH probe in the flocculation tank provides feedback control of neutralization. The coagulation-flocculation process enhances removal of colloidal and suspended solids in the subsequent primary sedimentation process. Flocculation tank effluent enters the five circular primary clarifiers for solids removal; clarifier effluent is then combined with return activated sludge for distribution to the three diffused air

FIGURE IV-1

SITE PLAN

EXISTING WASTEWATER TREATMENT FACILITIES



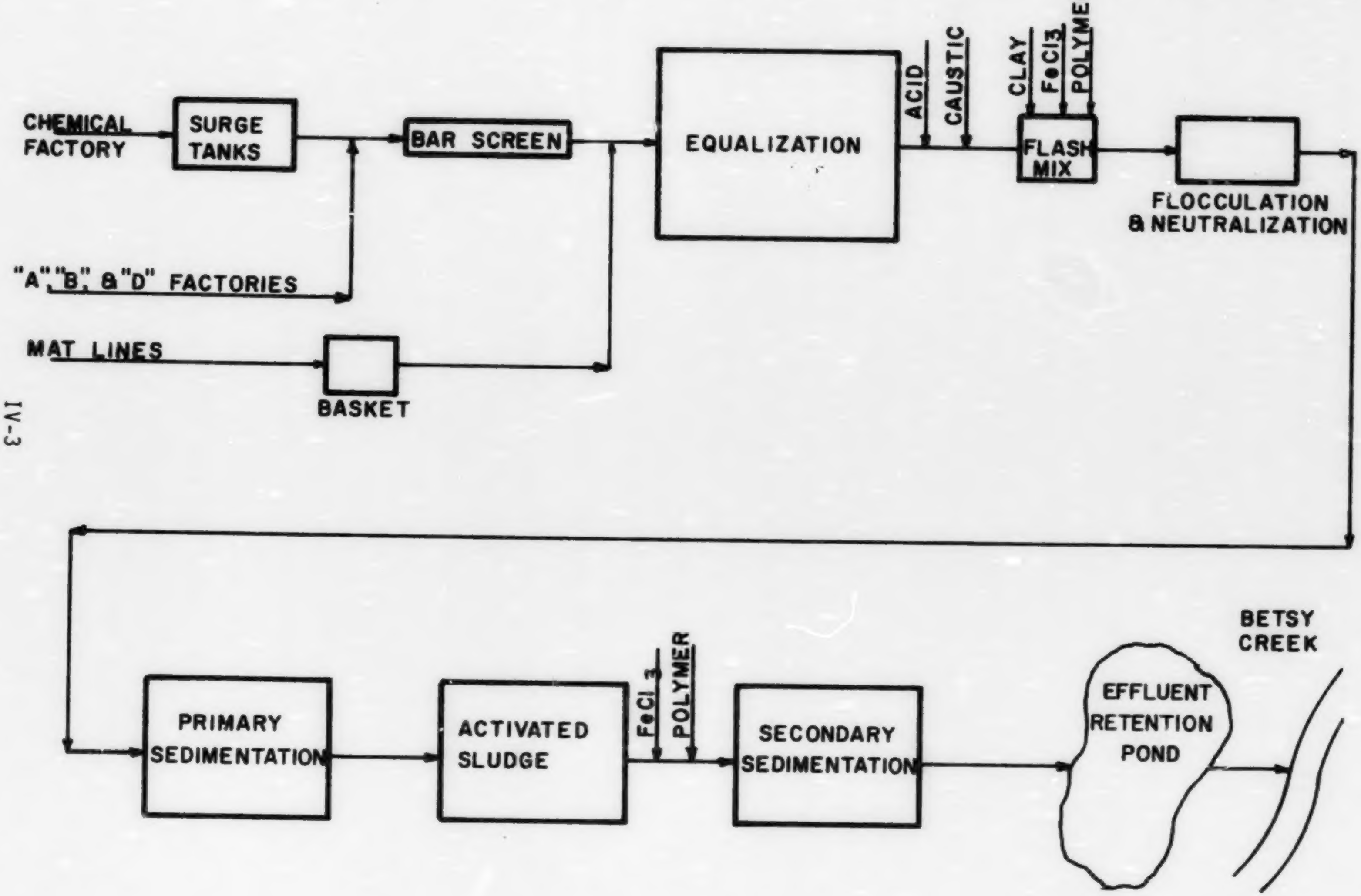


FIGURE IV-2
PROCESS FLOW DIAGRAM-EXISTING WASTEWATER TREATMENT FACILITIES

aeration basins. Mixed liquor from the aeration basins flows to the three rectangular secondary clarifiers. Ferric chloride and polymers are added prior to the clarifiers to enhance sedimentation efficiency. Clarified effluent flows through a Parshall flume and then is pumped to the Effluent Retention Pond. The pond discharge enters Betsy Creek. Sludge from the primary clarifiers is pumped into an aerobic digester from which it is periodically pumped to the sludge lagoon. Waste activated sludge is pumped to the flocculation tank and settles in the primary clarifiers. The treatment process unit volumes are given in Table IV-1.

EPA INVESTIGATIONS OF BIOLOGICAL TREATMENT OPERATIONS

In response to requests from OCF corporate environmental personnel, staff members of EPA's National Field Investigations Center worked in Anderson during March, April, and May, 1973, to develop operational control techniques to improve the performance of the wastewater treatment facility. The study concentrated on delineating process parameter responses and sludge wastage quantities. Recommendations made by EPA at the conclusion of the project are listed below:

- Use process demands to determine return sludge and waste sludge requirements.
- Install a recording-totalizing waste sludge flow meter.
- Provide improved screening and grit removal equipment.
- Improve the pH adjustment system, particularly the caustic feeding part of the system.
- Resolve the problem of handling highly concentrated organic batch dumps.
- Problem wastes should be categorized. These should include the highly concentrated organic wastes and those with extremely high or low pH values.
- Modify the blower system so that it can provide sufficient air under changing conditions.
- Consider using a recording D. O. analyzer in the aeration tanks.

TABLE IV-1

EXISTING TREATMENT PROCESS UNITS

PROCESS UNIT	DIMENSIONS	VOLUME (gal)
Chemical Wastewater Surge Tanks (3)	-	2 @ 22,500 1 @ 10,000
Large Equalization Basin	50' ϕ x 17'	249,679*
Small Equalization Basin	34' ϕ x 15'	101,869
Flash Mix Chamber	-	1,900
Flocculation Tank	-	11,000
Primary Clarifier No. 6	14' ϕ x 9'	10,363
Primary Clarifier No. 7	14' ϕ x 9'	10,363
Primary Clarifier No. 8	14' ϕ x 9'	10,363
Primary Clarifier No. 9	14' ϕ x 9'	10,363
Primary Clarifier No. 10	14' ϕ x 9'	10,363
Aeration Basin No. 1	50' x 25' x 15'	140,250
Aeration Basin No. 2	50' x 25' x 15'	140,250
Aeration Basin No. 3	50' x 25' x 15'	140,250
Secondary Clarifier No. 1	50' x 10' x 8.5'	31,790
Secondary Clarifier No. 2	50' x 10' x 8.5'	31,790
Secondary Clarifier No. 3	50' x 10' x 8.5'	31,790
Aerobic Digester	45' ϕ x 25'	297,411

*Maximum volume - water level varies considerably.

- Improve the distribution box between the aeration tanks and the clarifiers so that mixed liquor can be distributed to the clarifiers without pulsing and air entrainment.
- Use secondary clarifiers as required by plant flow rates.
- Reevaluate the secondary clarifier inlet piping.
- Consider an improved ferric chloride feed system.
- Consider installing additional polymer feed facilities.
- Improve the nutrient addition system to give the operator the ability to adjust and maintain required feed rates.

These recommendations either have been or will be implemented, as is discussed in a subsequent section of this chapter.

EVALUATION OF WASTEWATER TREATMENT FACILITIES

Dye studies of the major wastewater treatment units were performed to evaluate the existing hydraulic characteristics and to quantify undesirable hydraulic problems. Studies were performed on the equalization basins, the primary clarifiers, and the secondary clarifiers.

The basic procedure consisted of adding a measured amount of flourescent dye to the influent of the particular unit being evaluated and measuring the concentration of dye in the effluent as a function of time. In all cases, Rhodamine B-WT Dye was used as a tracer substance. The flow characteristics in the unit can be ascertained from the shape of the dye recovery curve.

The effluent concentration of dye was measured with a Turner Fluorometer Model No. 111 equipped with a 546 primary filter and a 590 secondary filter. The fluorometer was calibrated using serial dilutions of the respective dyes at 25°C. All samples were brought to the calibration temperature before measurement.

The basic purpose of the studies was to determine the relative amounts of mixing, plug flow, and dead space occurring in each unit process and to compare the actual results with the desirable characteristics.

Complex mathematical models have been derived for describing the various combinations of flow characteristics that occur in a theoretical hydraulic system. Applications of the theoretical models to real systems have, in some cases, been quite satisfactory. The disadvantage of using complex models, however, is that the original purpose of the flow study can be lost in the complexity of the analysis.

The method utilized in this study is based on flow models that can be presented graphically as shown in Figures IV-3 through IV-5. Figure IV-3 depicts the effect of dead space on a completely mixed flow system. It can be shown theoretically that approximately 63% of the dye added to a completely mixed system will be recovered after one detention time:

$$\text{Percent Recovery} = 100\ (1 - e^{-t/T}) = 100\ (1 - e^{-1.0}) = 63 \qquad \text{(IV-1)}$$

where:

T = detention time

t = actual measured time interval

FIGURE IV-3

THEORETICAL DYE RECOVERY CURVES FOR
A COMPLETELY MIXED SYSTEM WITH VARYING
AMOUNTS OF DEAD SPACE

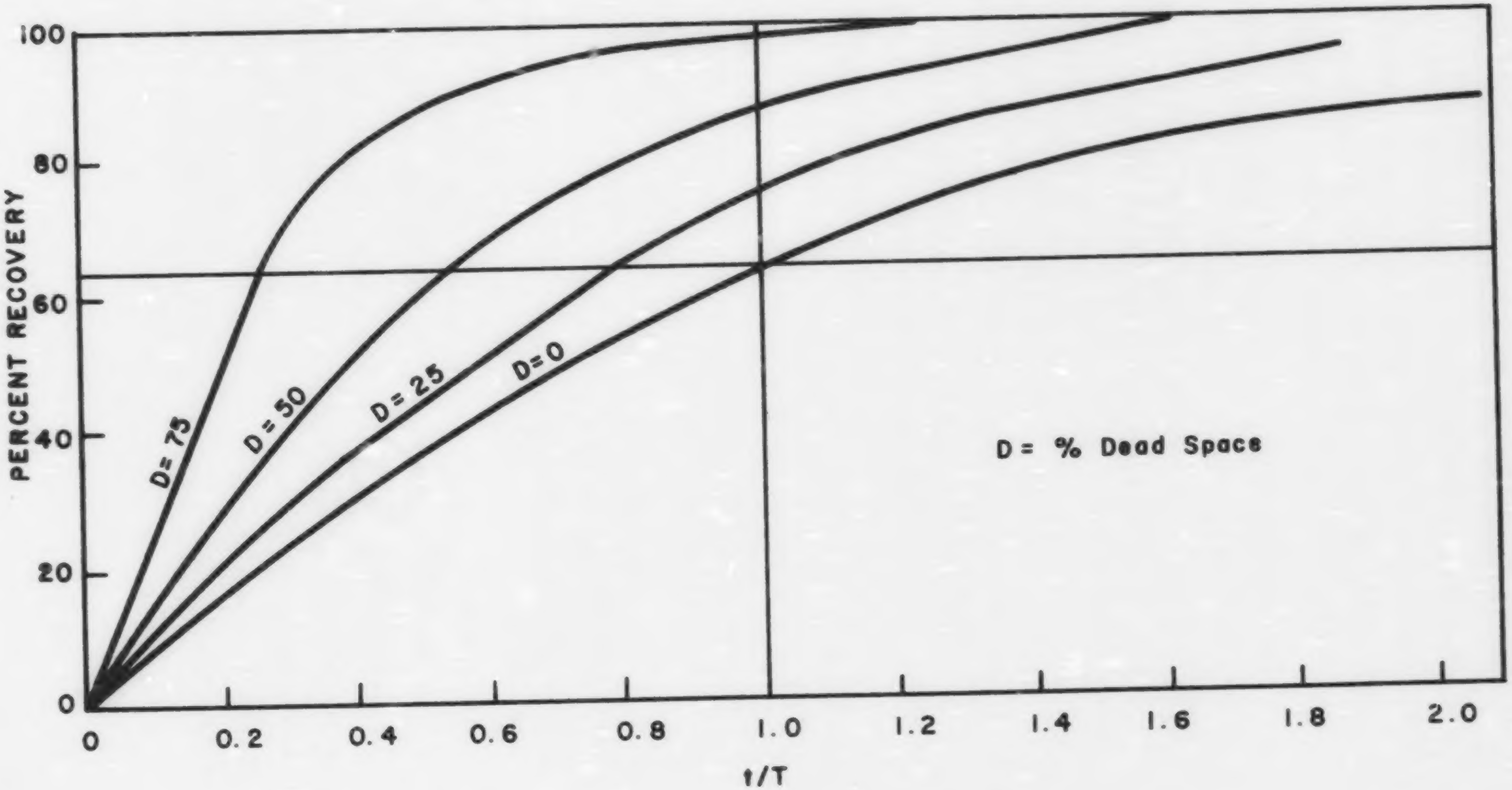


FIGURE IV-4

THEORETICAL DYE RECOVERY CURVES FOR
A COMPLETELY MIXED SYSTEM WITH VARYING
AMOUNTS OF PLUG FLOW

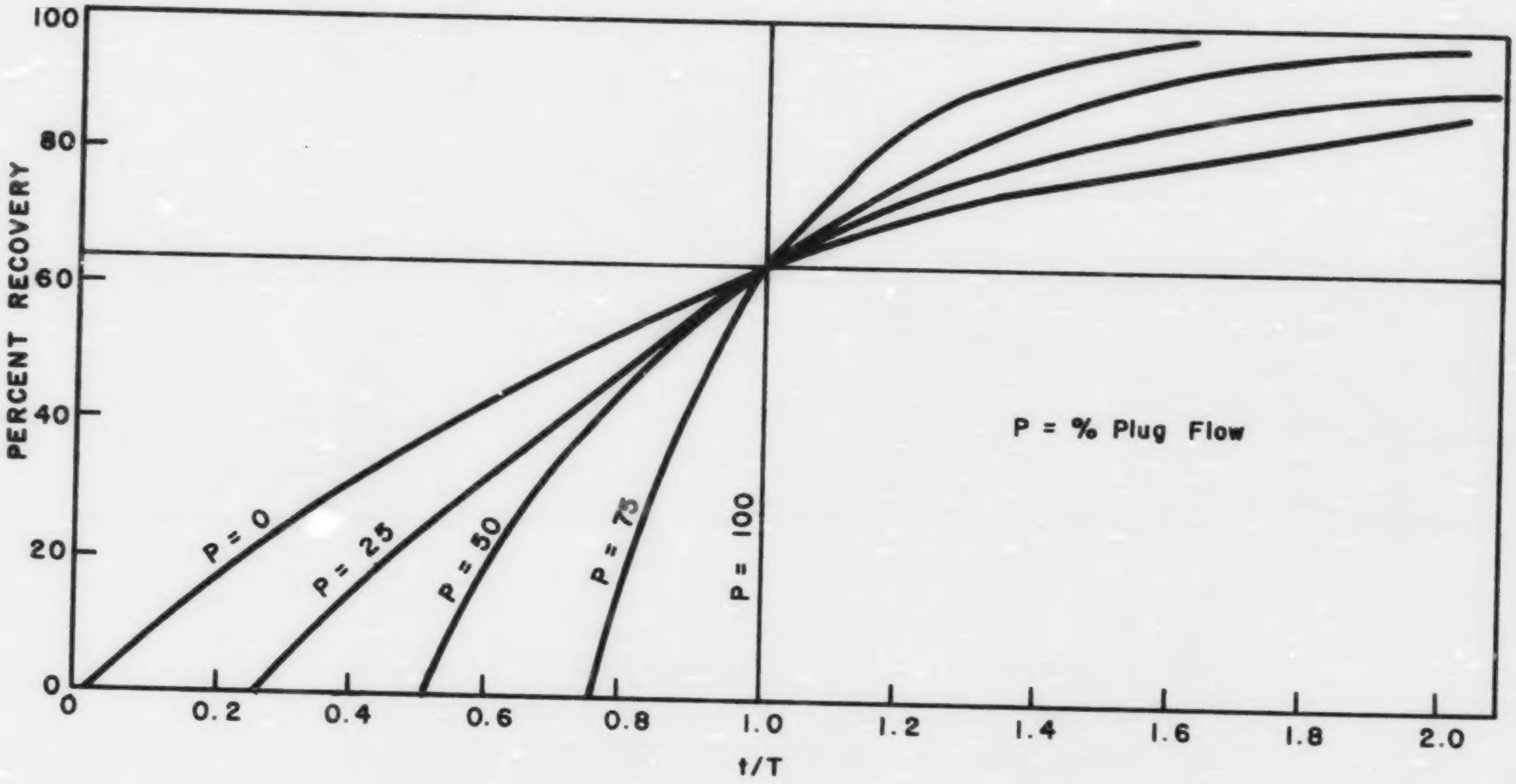
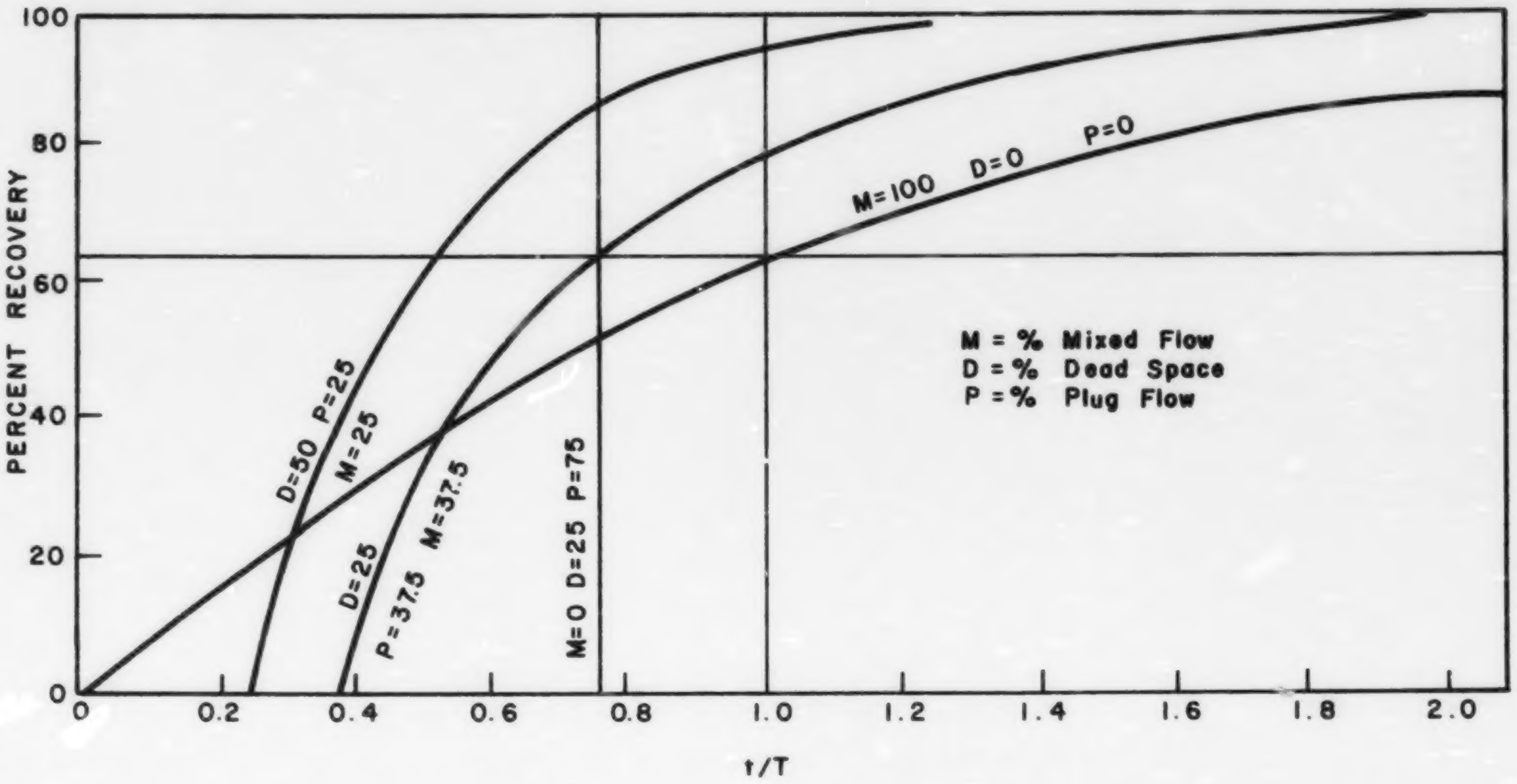


FIGURE - IV-5

THEORETICAL DYE RECOVERY CURVES FOR
A COMPLETELY MIXED SYSTEM WITH VARYING
AMOUNTS OF DEAD SPACE AND PLUG FLOW



To determine the amount of dead space in a completely mixed system, it is necessary only to determine at what fraction of a detention time 63% of the dye is recovered. The remaining fraction is then equal to the dead space in the vessel. As shown in Figure IV-3, if 63% of the dye is recovered at t/T = 0.75, the amount of dead space is 0.25 or 25%.

Figure IV-4 shows the effect of plug flow on a completely mixed system. In this case, all of the curves pass through 63% dye recovery at t/T = 1.0, but the curves originate at various points on the abscissa. The fraction of plug flow is equal to the starting point on the abscissa. For example, a completely mixed system having 50% plug flow would have a recovery curve originating at t/T = 0.50 as shown in Figure IV-4.

Figure IV-5 shows various combinations of plug flow and dead space in a completely mixed system. The determination of the relative amounts of the three characteristics proceeds exactly as for the individual cases. For example, a system having 25% dead space would show 63% dye recovery at t/T = 0.75. If in the same system the remaining volume - i.e., the effective volume - were divided evenly between completely mixed and plug flow, the curve would originate at t/T = 0.375 - i.e., one half of (1.0 - 0.25).

The concentration of dye in the effluent from a particular unit was measured versus time. This was then plotted with concentration as the ordinate and time as the abscissa. For ease of analysis, both parameters are "normalized" by dividing the concentration C by C_o and the time t by T.

where:

C and t = actual concentration of dye after a particular time interval t

$$C_o = \frac{\text{weight of dye added}}{\text{theoretical volume of tank}}$$

$$T = \text{detention time} = \frac{\text{volume}}{\text{flow}}$$

The actual percent recovery of dye is then equal to the area under the curve of C/C_o versus t/T. A percent recovery versus time curve can be established by integrating the curve for various time intervals. The latter curve is constructed by assuming that the area under the concentration versus time curve is equal to 100% dye recovery rather than the actual dye recovery.

Dye recovery curves for each of the unit processes investigated are presented in Figures IV-6 through IV-12. Two different curves are presented in each figure: the left ordinate refers to the normalized concentration C/C_o and the right ordinate refers to the percent recovery assuming 100% dye recovery at $t/T = \infty$. For both curves, the abscissa is the normalized time interval as t/T. The hydraulic characteristics of the units in terms of the relative percent of dead space, plug flow, and completely mixed flow are stated on each figure and summarized in Table IV-2.

Equalization Basins

The hydraulic studies performed on the equalization basins entailed the injection of dye at both of the wastewater entrance points, namely the Mat Line effluent and the bar screen effluent. Amounts of dye proportional to the flow at each point were released simultaneously. The combined effluent of the equalization basins was then monitored such that the total equalization facilities were treated as one system. The results of the hydraulic study on the equalization system indicate that approximately 78 % of the system is completely mixed, 4% is plug flow, and 18% is dead space (Figure IV-6). On the basis of these results, the equalization capacity of the total system is approximately equal to the operating volume of the system since the flow regime is primarily completely mixed. It is suspected that some, if not all, of the dead space reflected in the dye study is due to solids deposition in the bottom of each of the equalization basins.

Primary Clarifiers No. 8, 9, and 10

Hydraulic studies were completed on the three most western primary clarifiers (Nos. 8, 9, and 10). The results of these dye studies as shown in Figures IV-7 through IV-9 indicate that each clarifier has basically the same hydraulic characteristics, although Clarifier No. 8 has 13% dead space while the others had none. This may have been due to the quantity of settled sludge present in the clarifier at the time of the test. Stored sludge would represent dead space and

TABLE IV-2

SUMMARY OF DYE STUDY RESULTS AND FLOW CHARACTERISTICS

Process Unit	Figure	Mean Flow Rate (gpm)	Flow Characteristics		
			Completely Mixed (percent)	Plug Flow (percent)	Dead Space (percent)
Equalization Basins	IV-6	350	78	4	18
Primary Clarifier No. 8	IV-7	60	79	8	13
Primary Clarifier No. 9	IV-8	84	92	8	0
Primary Clarifier No. 10	IV-9	75	90	10	0
Final Clarifier No. 1	IV-10	113	90	10	0
Final Clarifier No. 2	IV-11	150	79	7	14
Final Clarifier No. 3*	IV-12	101	64	7	29

*After target plate system re-installed.

FIGURE IV-6

DYE STUDY FOR EQUALIZATION BASINS

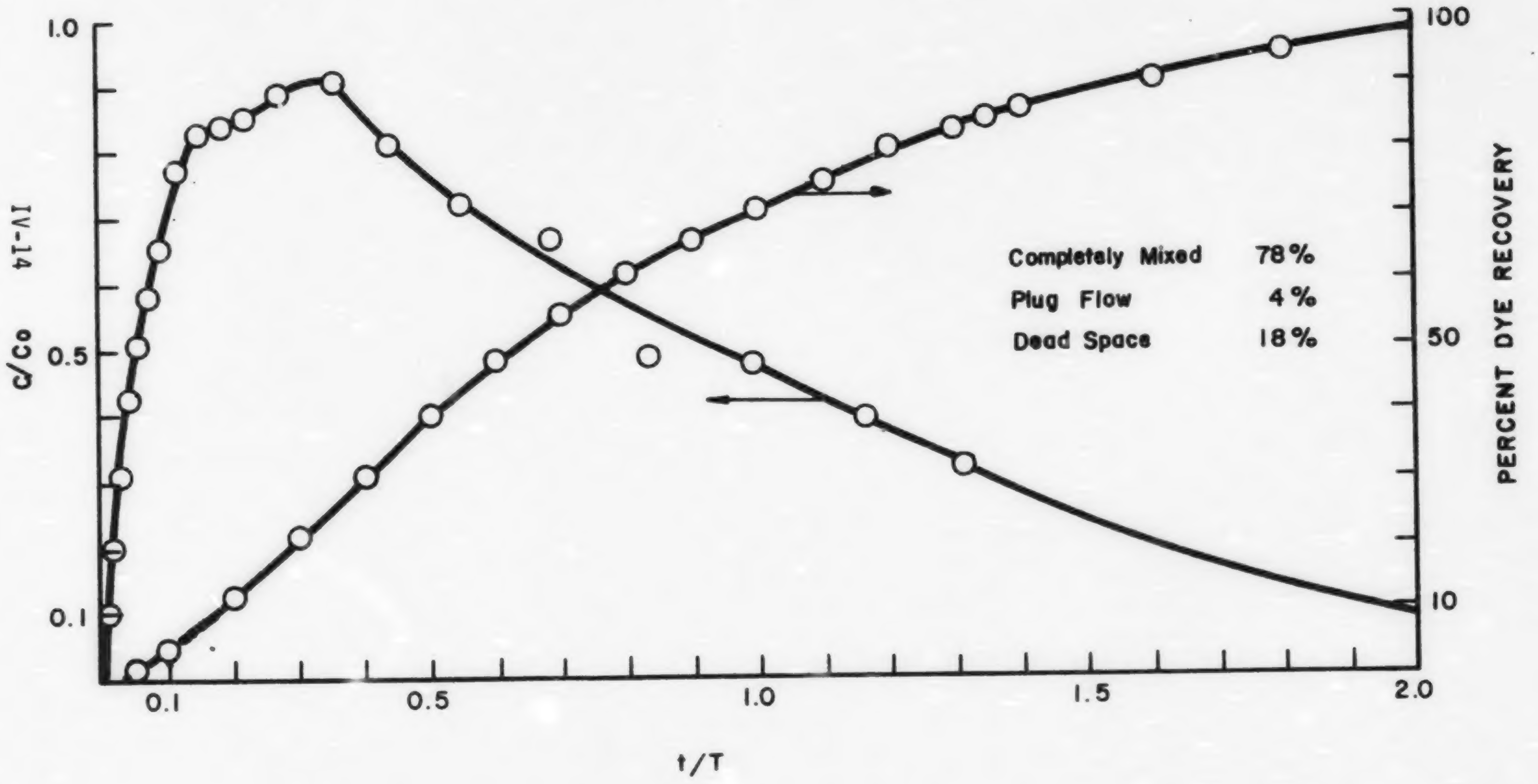


FIGURE IV-7

DYE STUDY FOR PRIMARY CLARIFIER #8

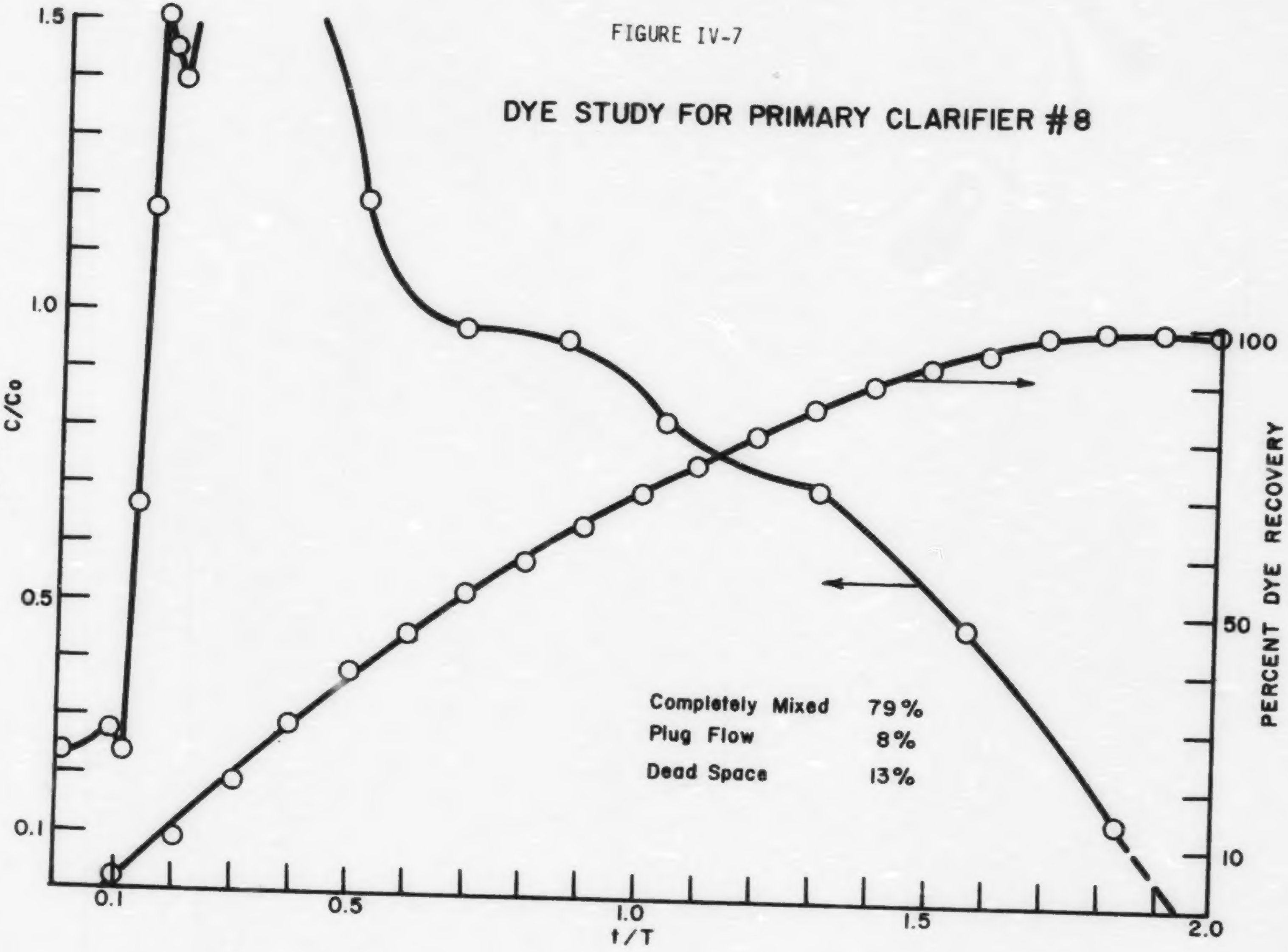


FIGURE VI-8

DYE STUDY FOR PRIMARY CLARIFIER #9

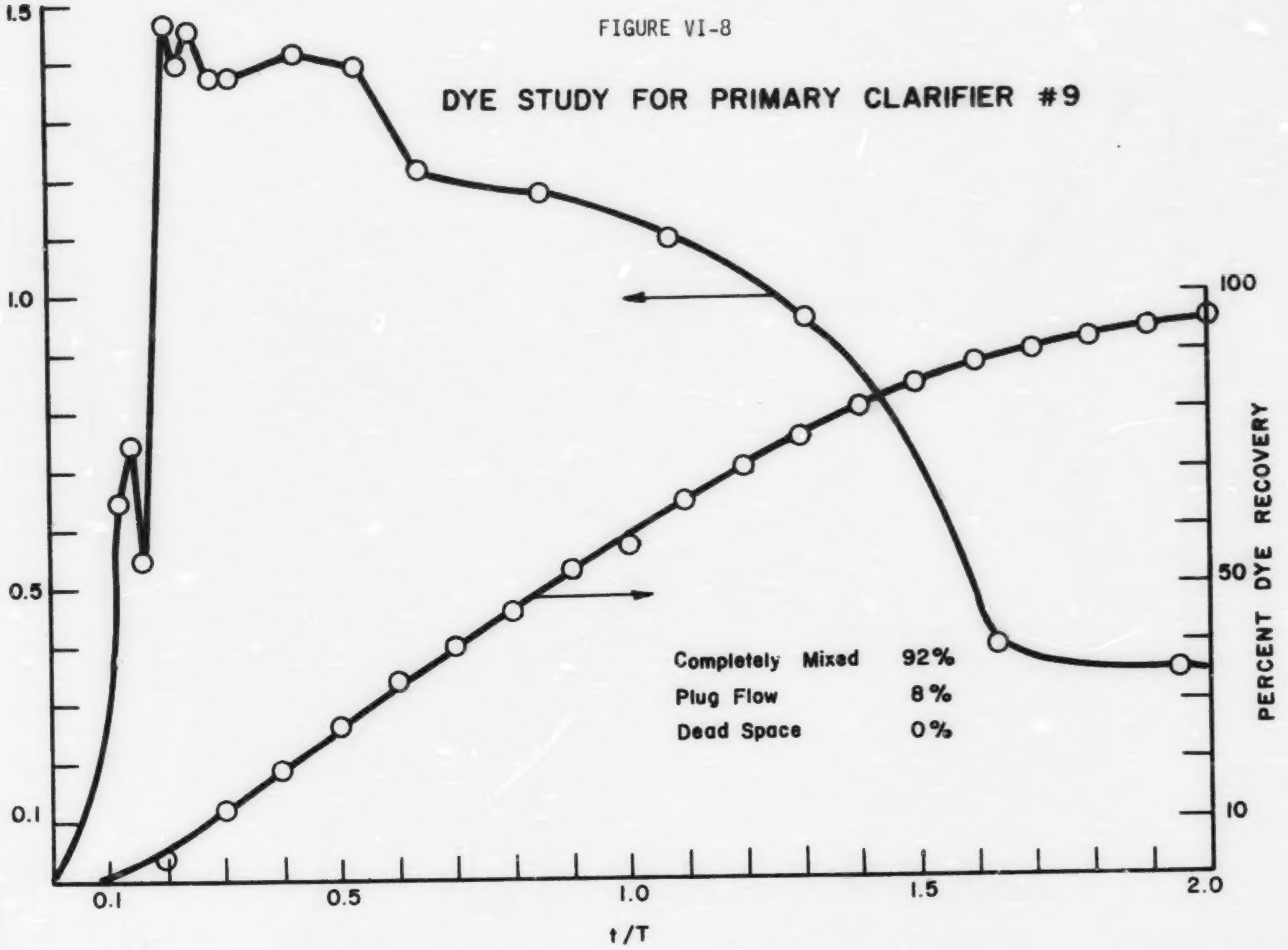


FIGURE IV-9

DYE STUDY FOR PRIMARY CLARIFIER #10

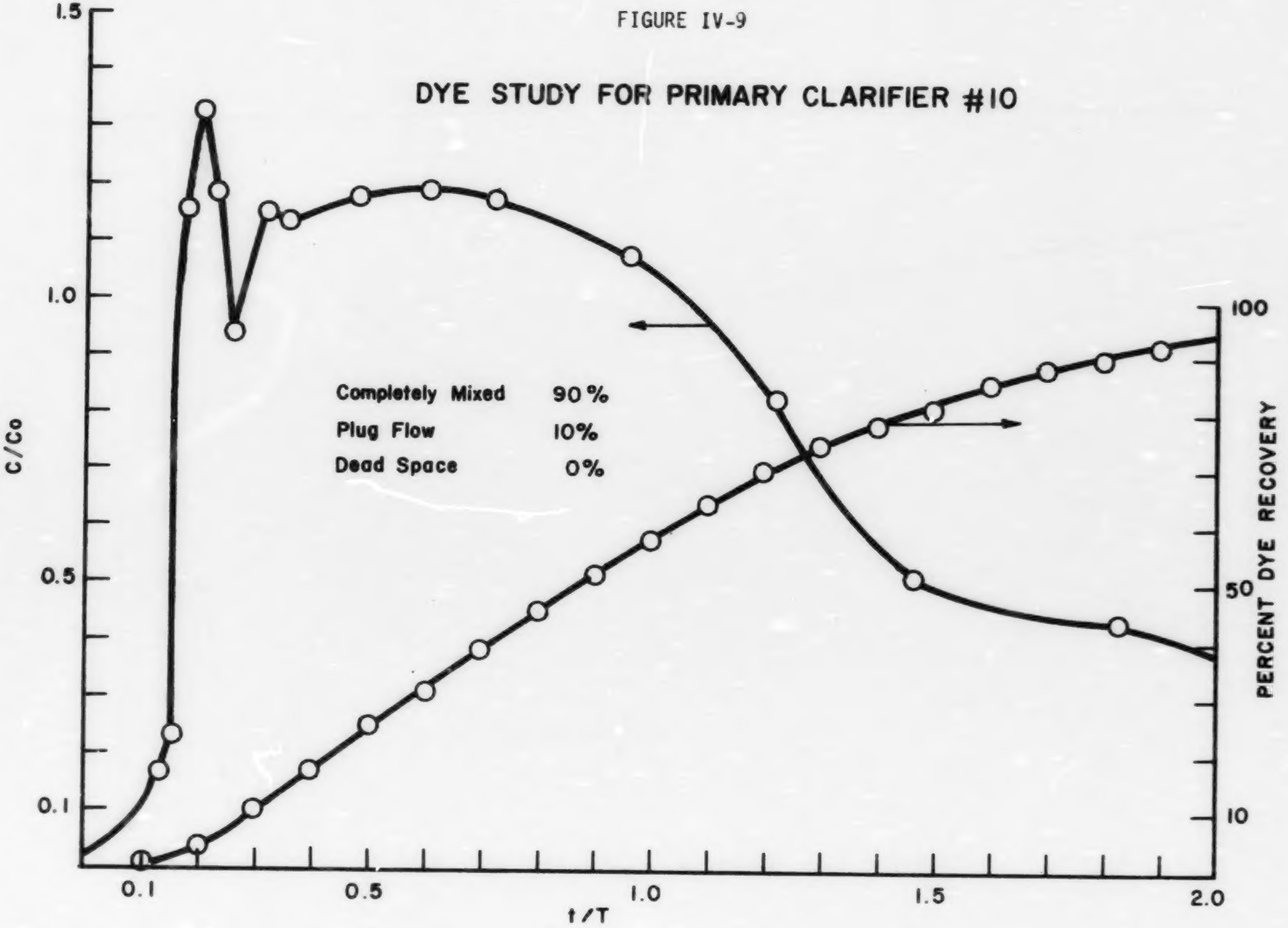
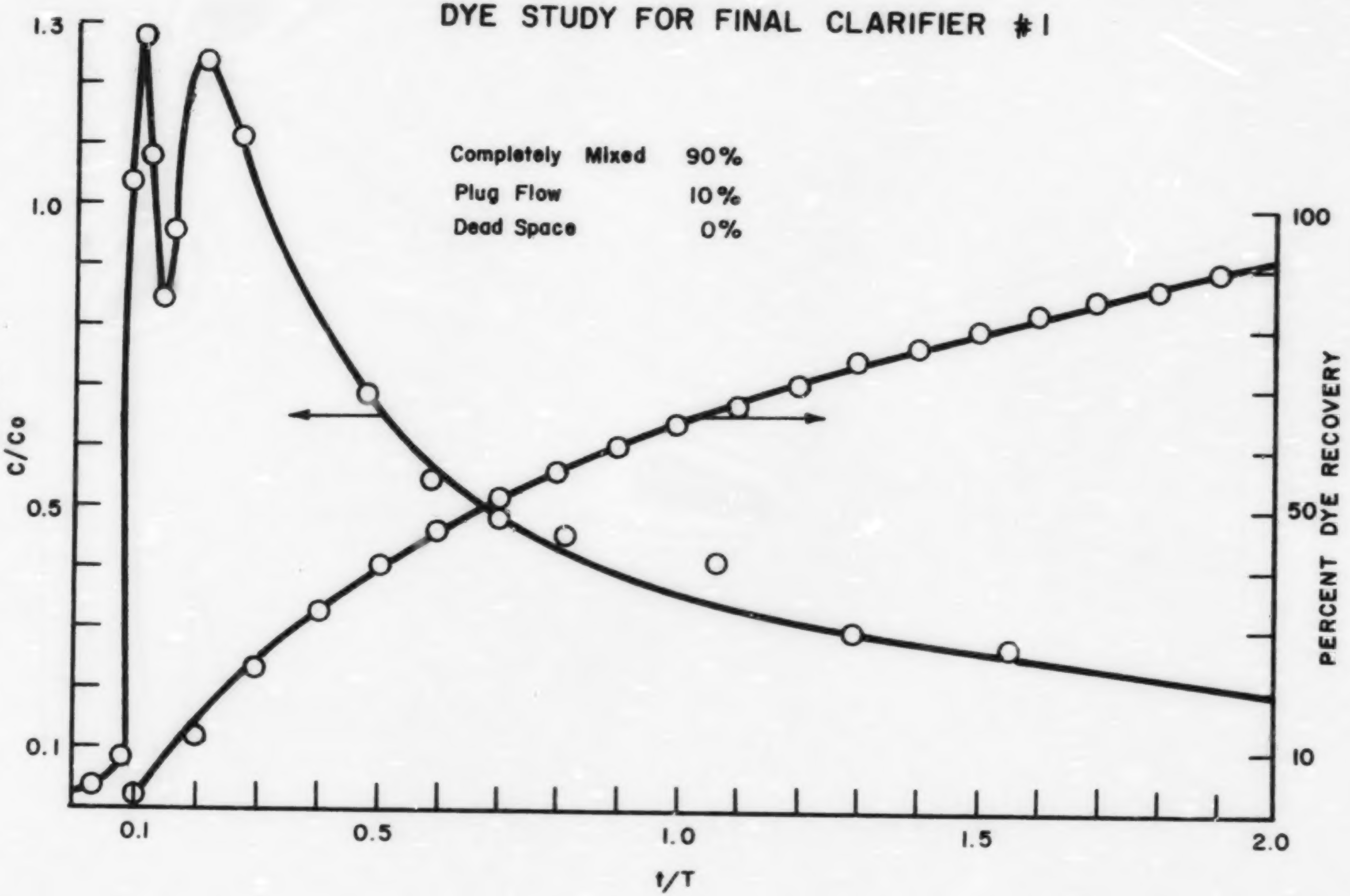


FIGURE IV-10



DYE STUDY FOR FINAL CLARIFIER #1

FIGURE IV-11

DYE STUDY FOR FINAL CLARIFIER #2

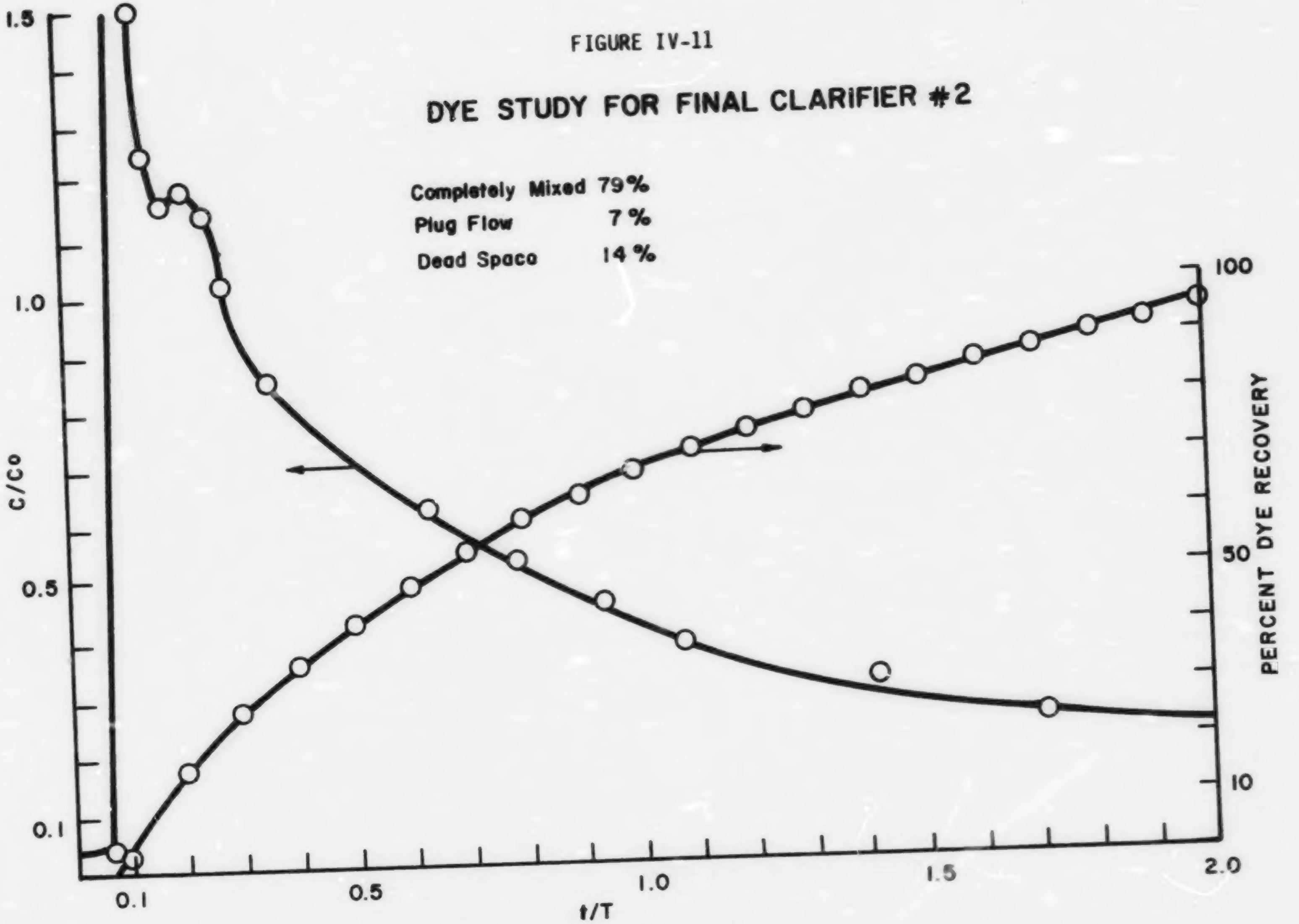
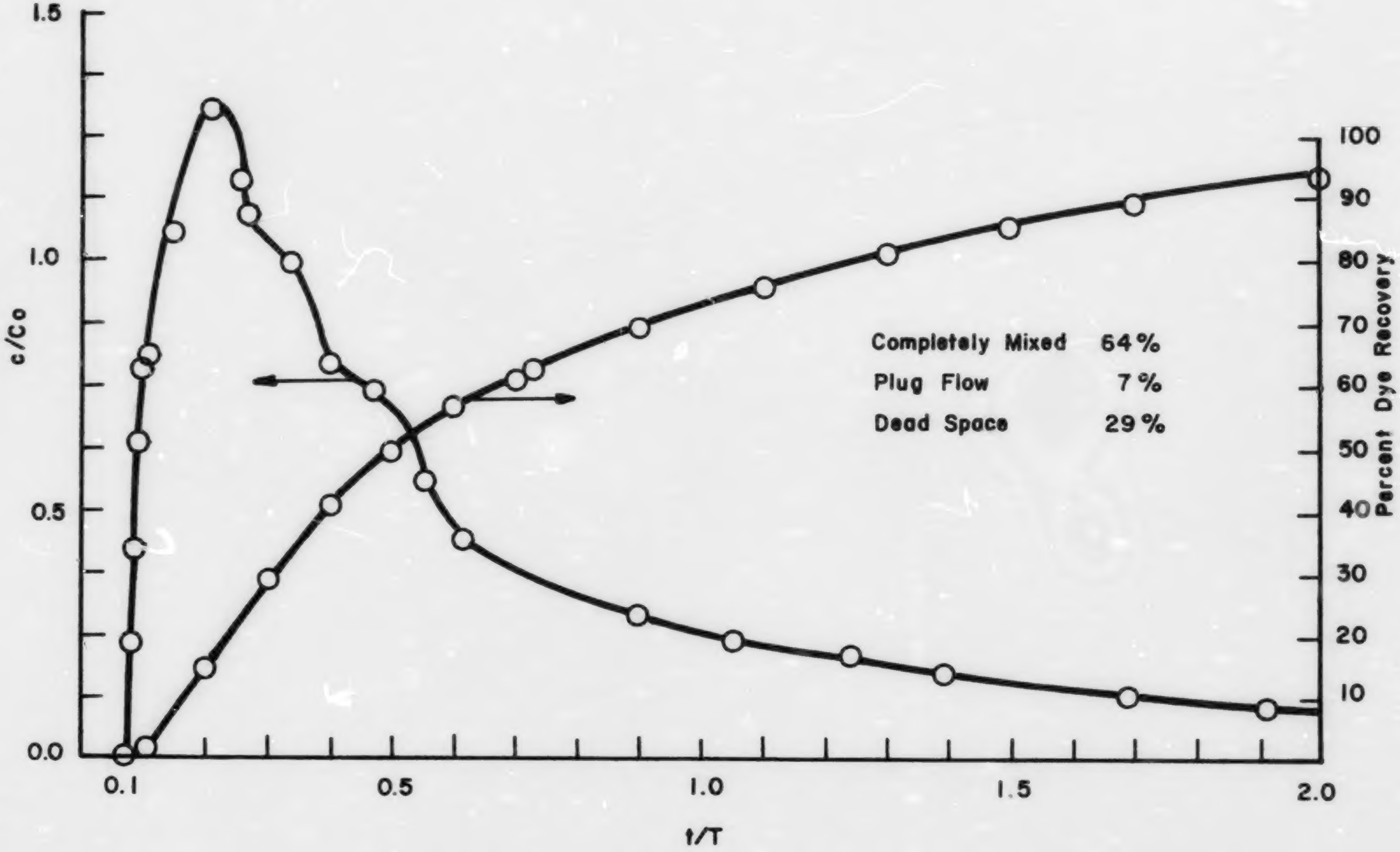

Completely Mixed 79%
Plug Flow 7%
Dead Spaco 14%

FIGURE IV-12

DYE STUDY FOR FINAL CLARIFIER # 3
FOLLOWING INTAKE MODIFICATION



as sludge is not withdrawn continuously, this could account for the difference. In general, while it appears that the hydraulic characteristics of the primary clarifiers are not within the acceptable operating range (20% plug flow), their performance is excellent.

Final Clarifiers No. 1 and 2

The hydraulic studies of the two final clarifiers (No. 1 and 2) entailed the measurement of the effluent concentrations as shown in Figures IV-10 and IV-11. With respect to the effluent measurements, each clarifier has relatively high completely mixed characteristics (90% and 79%) and relatively low plug flow characteristics (10% and 7%), which are not favorable for clarification systems. Visual observations made during the studies indicate that turbulence patterns caused by the downward entrance of the influent liquid into the clarifiers are largely responsible for an observed boiling effect at approximately one-third of the length of the clarifier. This situation is not conducive to proper clarification or thickening of the activated sludge.

Modified Final Clarifier - No. 3

Based on these studies the secondary clarifier influent distribution system was modified to eliminate the vertical velocities introduced by the surface release of the mixed liquor. In an attempt to accomplish this, the inlet configuration for Final Clarifier No. 3 was returned to the inlet pipe and target plate system originally designed. The dye study conducted after the modification gave results as shown in Figure IV-12.

The results indicate that no additional percentage of plug flow resulted from this modification. However, considerably more dead space was recorded (29%). This probably would not help nor hinder the clarification or thickening capability of the clarifier, but does reduce the volume and surface area utilized. The loss is not critical, however, as the surface overflow rates on these clarifiers are already fairly conservative. Visually, the boiling effect one-third down the length of the clarifier disappeared. However, in order to improve the hydraulic characteristics of all 3 clarifiers (i.e., increase plug flow) a new influent distribution system should be designed. Flow should be introduced into the clarifier over the entire width of the tank. Baffles may be used to dissipate inlet energy and thus provide a greater area for quiescent settling. Improvement measures of this nature will increase sedimentation efficiency.

RAW WASTE LOAD REDUCTIONS

Results of the 1973-1974 wastewater survey and characterization study indicated that the majority of the wastewater flow and contamination originated in the fibrous glass production areas of "A" Factory and "D" Factory. In order to improve efficiencies of the treatment units, recommendations were made as to possible reductions in the raw waste load entering the treatment plant. Reductions in both binder loss and wastewater flow were suggested.

Wastewater contaminants consist almost entirely of either direct or indirect binder losses. Water usage and wastewater treatment costs, especially for a recirculation system, can be reduced considerably through diminution of process wastewater flows. Candidate streams for flow reduction include:

- Scrubber blowdown - by metering or locking makeup valves so as to allow the minimum required water use; and
- Washdowns - through installation of self-closing valves on washdown hoses.

Flow reduction efforts will be discussed more completely in another section.

TREATMENT FACILITY IMPROVEMENTS

Wastewater treatment facilities at the Anderson Plant have been modified considerably during the last several years, not only in terms of physical changes but also operational changes. Furthermore, numerous improvements will be made in 1976 and 1977. Recent treatment facility improvements are discussed in this section.

The operational control procedures demonstrated by EPA during their technical support project in 1973 are still being used at the plant to determine return sludge and waste sludge quantities. A recording-totalizing flow meter has been installed on the waste sludge line to quantify wasteage volumes. The blower system capacity has been increased to three times its original value in order to constantly maintain aerobic conditions in the aeration basins. This was supplemented by installation of a recording dissolved oxygen (D.O.) analyzer to monitor changes in D.O. concentration. The flow-splitting system between the aeration basins and the final clarifiers has been improved to eliminate pulsation and air entrainment. Nutrient storage tanks have been raised above the gound surface to enhance preventive maintenance procedures.

Based upon treatability studies conducted in 1974, temporary facilities for coagulant addition (to the primary clarifier influent wastewater) were constructed and became operational in May, 1975. Types and quantities of coagulant addition are as follows:

Clay (Nalco 8151)................................	166 mg/l
Ferric Chloride (40% $FeCl_3$)......................	45 mg/l
Anionic Polymer (American Cyanamid 837A)..........	0.3 mg/l
Cationic Polymer (Hercules 834.1)................	12 mg/l

Sedimentation efficiency has been increased considerably, as will be discussed in the next section of this chapter. The three surge tanks (total volume = 55,000 gal) which receive Chemical Plant process wastewater became operational in September, 1975. The surge capability allows this particular wastewater stream to "bleed" into the equalization basins at a controlled rate, thereby eliminating shock organic loads.

HISTORICAL PERFORMANCE RECORD

Treatment performance data from January, 1972 to May, 1976 are used in this section to describe the historical record of treatment operations for the existing facilities. Monthly average values for BOD_5 COD, and TSS concentrations in the primary clarifier influent, primary clarifier effluent, and secondary clarifier effluent are depicted in Figures IV-13 to IV-15. During the past year (June, 1975 to May, 1976) the raw waste load from the manufacturing facility has decreased from the high levels occurring in 1973 and 1974 to those observed during 1972. At the same time, secondary clarifier effluent concentrations have decreased, as has wastewater flow rate (Figure IV-16B). Wastewater treatment facility performance is summarized in Table IV-3.

Monthly average removal efficiencies are shown in temporal plots as Figures IV-16A, 17A, 17B, 18A, and 18B. Coagulant addition practices have improved and stablized TSS removal efficiencies, as shown in Figures IV-16A and 18A. Overall facility performance during 1975 and 1976 is illustrated by the removal efficiencies listed below:

BOD_5	94% - 98%
COD	85% - 93%
TSS	95% - 99%
TOC	85% - 95%

Obviously, the treatment facility exhibits exceptional performance in comparison to similar industrial installations.

Correlations between wastewater BOD_5 and COD concentrations are depicted in Figure IV-19. The "r" values on the graph are correlation coefficients obtained from least-squares linear regression analyses, which resulted in development of the following relationships:

Primary Influent : $COD = 2.0\ (BOD_5) + 740$
Primary Effluent : $COD = 2.4\ (BOD_5) + 10$
Secondary Effluent : $COD = 2.7\ (BOD_5) + 140$

A similar analysis of BOD_5 and TOC concentrations resulted in the relationships listed below:

Primary Influent : $TOC = 0.53\ (BOD_5) + 260$
$r = 0.48$
Secondary Effluent : $TOC = 2.1\ (BOD_5) + 11$
$r = 0.98$

Prior to coagulant addition, secondary effluent BOD_5 and TSS concentrations were related in the following manner:

$$TSS = 0.30\ (BOD_5) + 14$$
$$r = 0.31$$

TABLE IV-3

PERFORMANCE SUMMARY

MEAN VALUES

YEAR	FLOW (gpm)	BOD$_5$[1]			COD[1]			TOC[1]			TSS[1]		
		PI[2]	PE[3]	SE[4]	PI	PE	SE	PI	PE	SE	PI	PE	SE
1972	335	369	-	14	1352	614	141	-	-	-	440	94	23
1973	374	450	348	18	1712	847	220	-	-	-	553	143	35
1974	333	528	-	19	1975	764	226	-	-	-	617	114	18
1975	296	313	-	11	1233	477	124	376	150	32	396	56	8
1976[5]	303	255	-	10	-	-	-	452	139	36	604	59	13

1. Concentration in terms of mg/l.
2. PI = primary clarifier influent.
3. PE = primary clarifier effluent.
4. SE = secondary clarifier effluent.
5. Values are for five months operation.

FIGURE IV-13

BOD$_5$ PERFORMANCE - MONTHLY AVERAGES

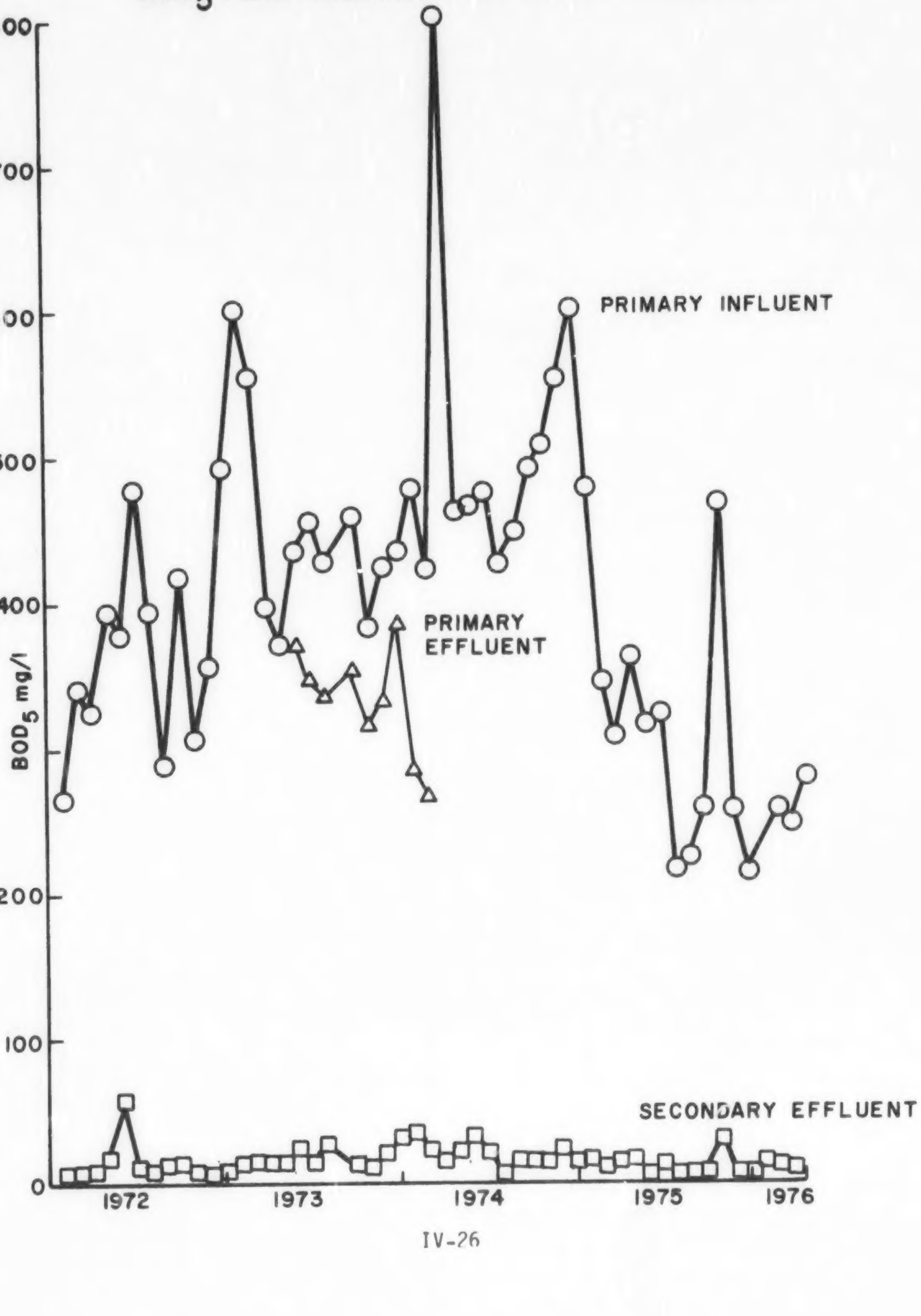
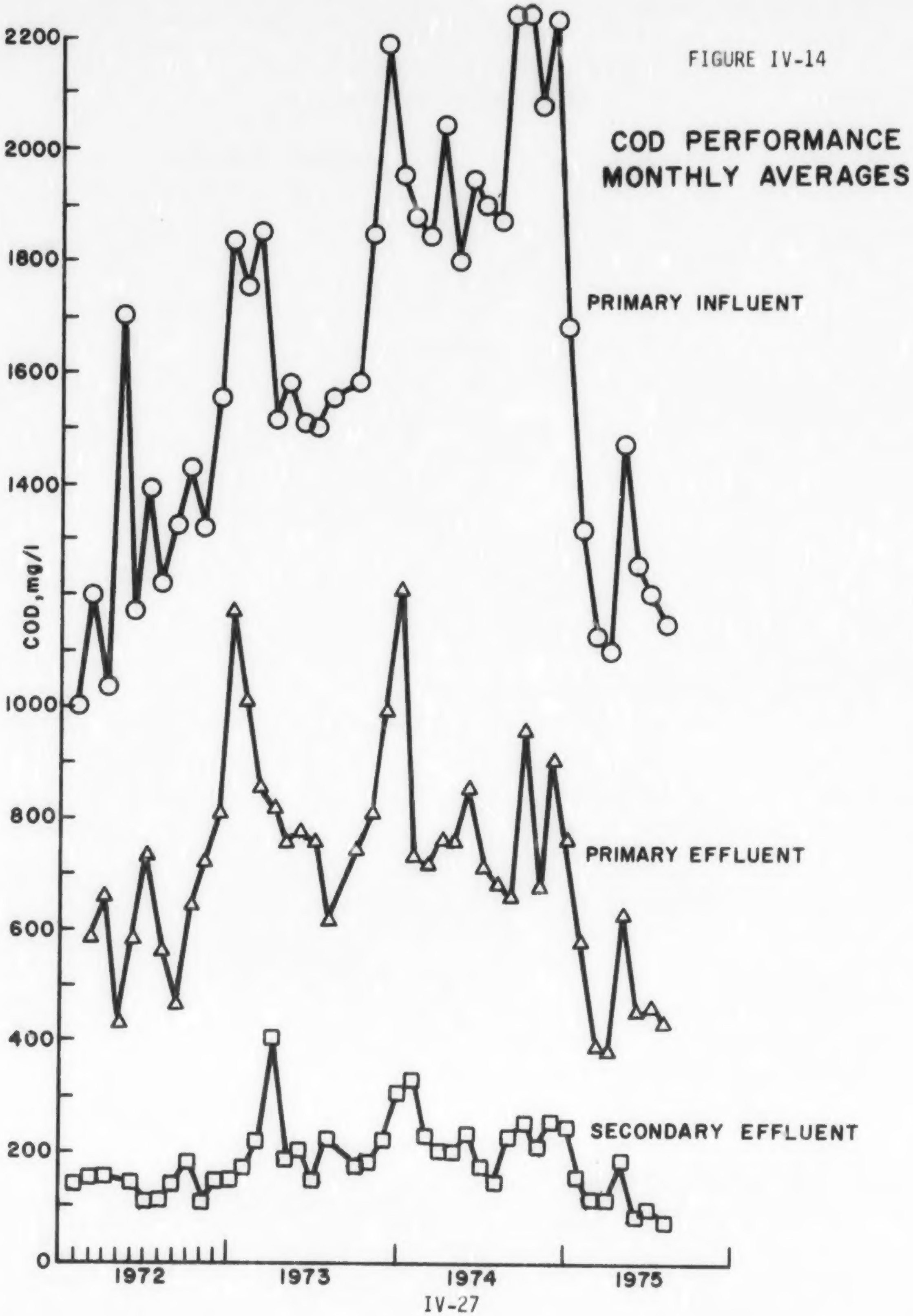


FIGURE IV-14



COD PERFORMANCE MONTHLY AVERAGES

FIGURE IV-15

TOTAL SUSPENDED SOLIDS
MONTHLY AVERAGES

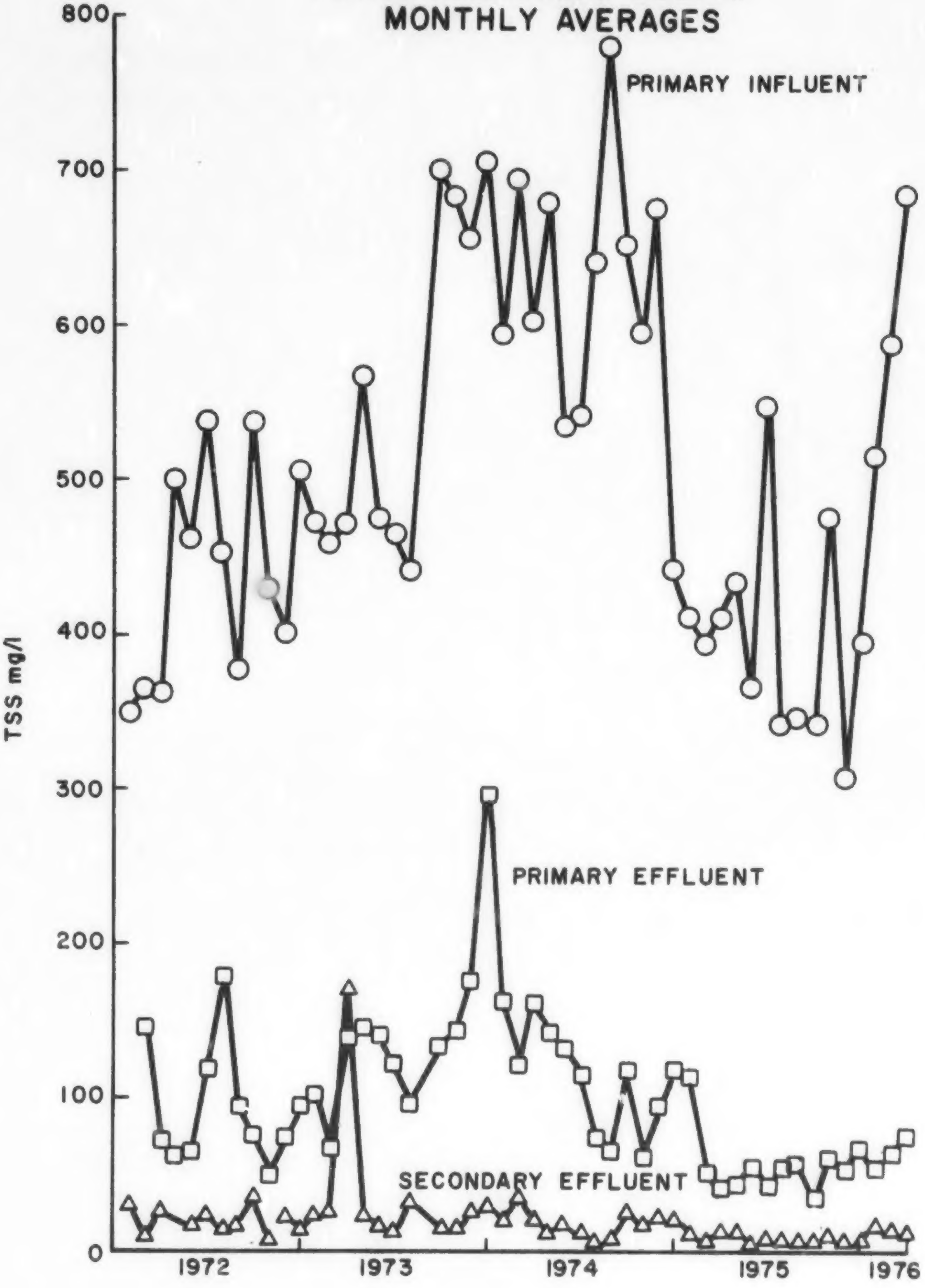


FIGURE IV-16A

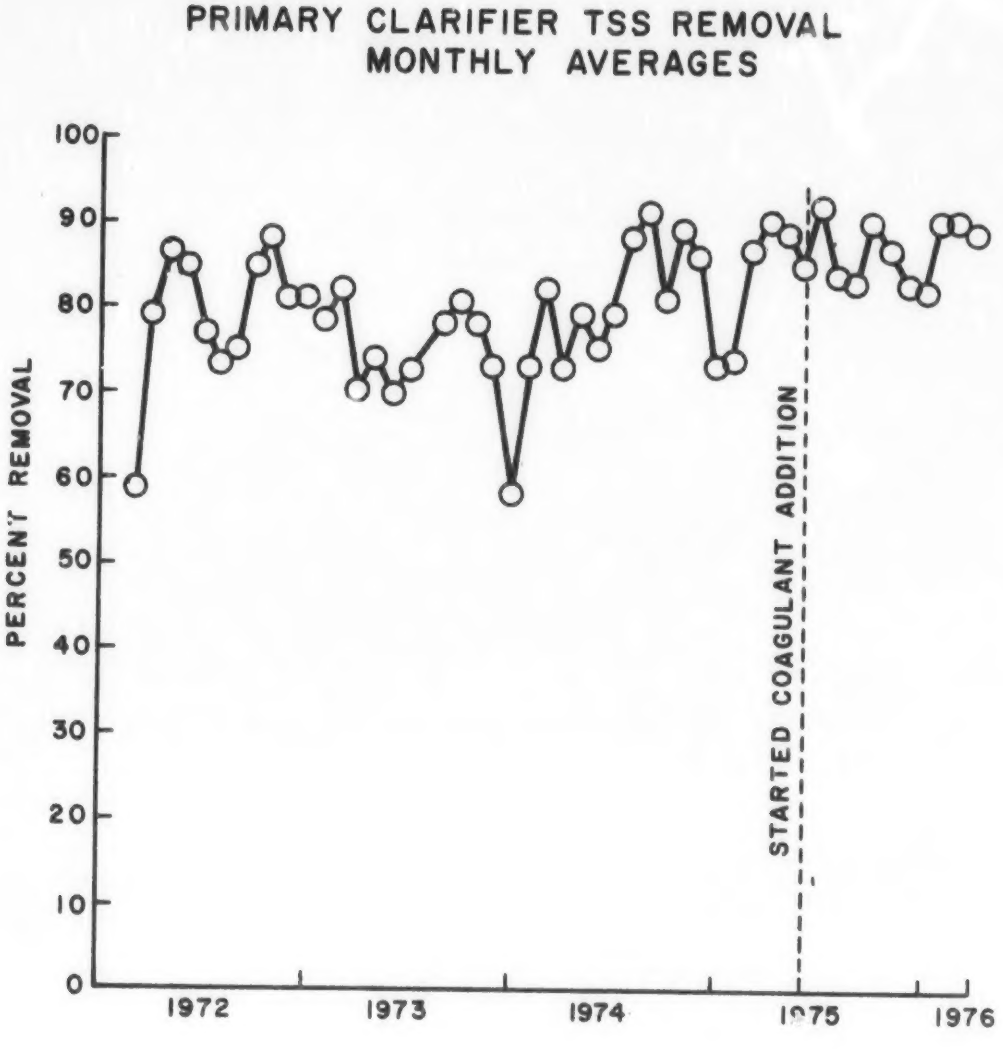


FIGURE IV-16B

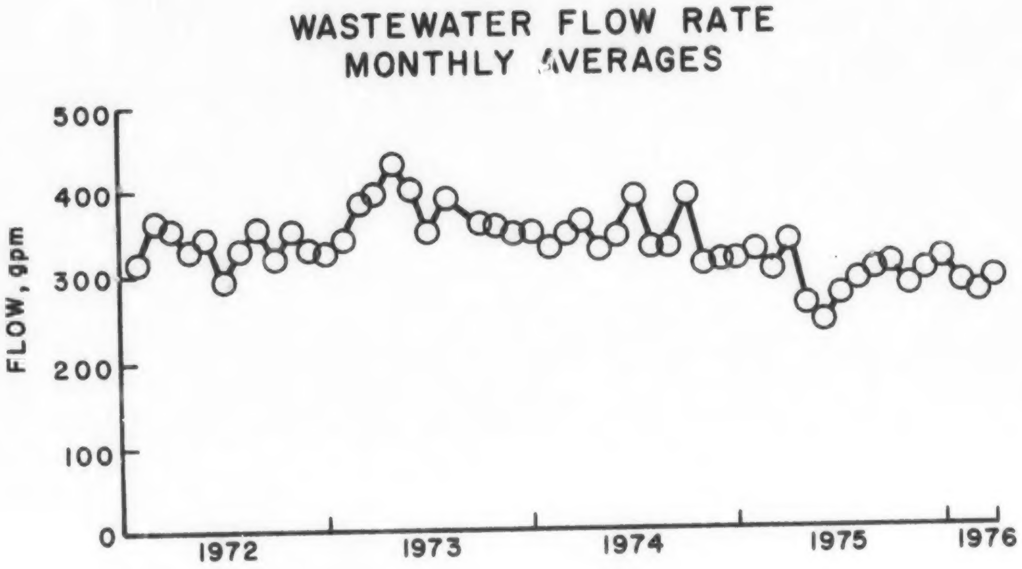


FIGURE IV-17A

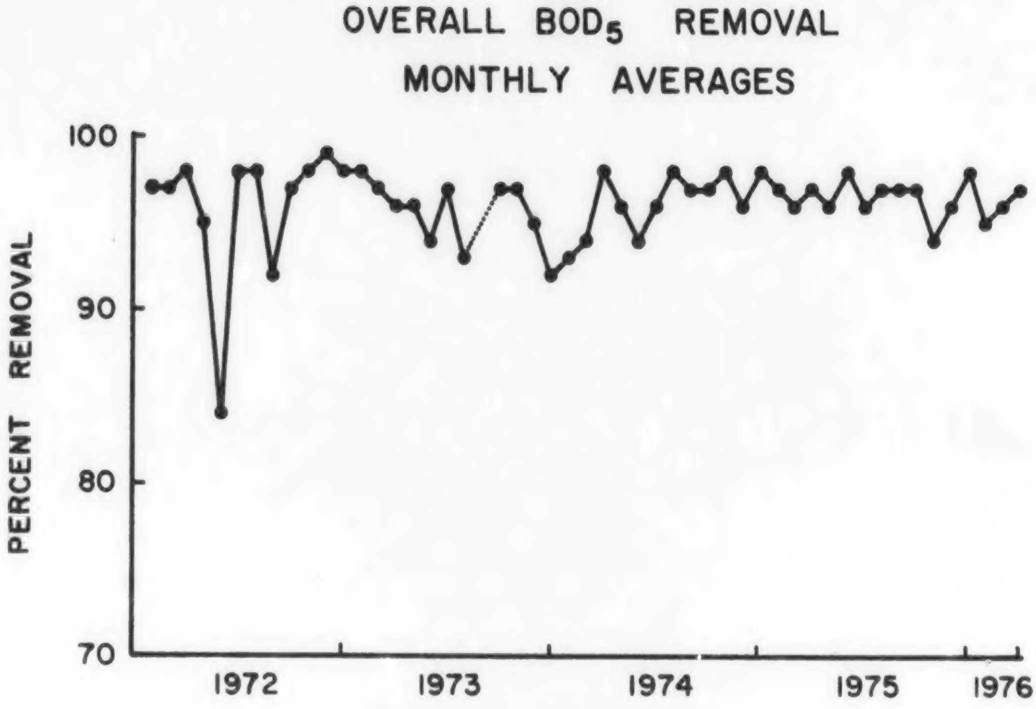


FIGURE IV-17B

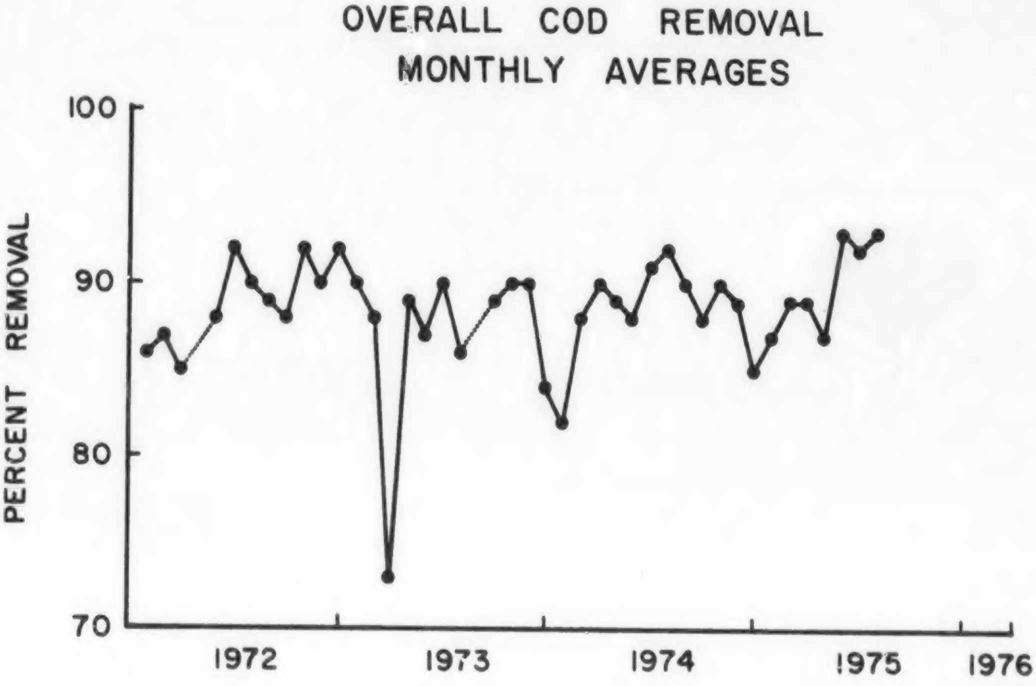


FIGURE IV-18A

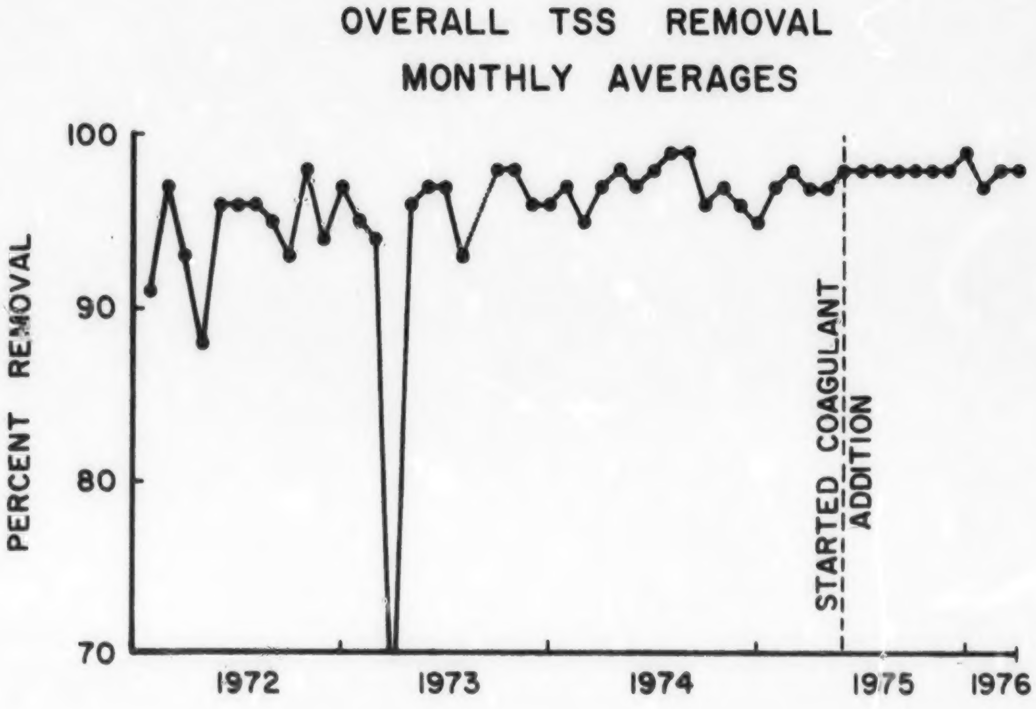
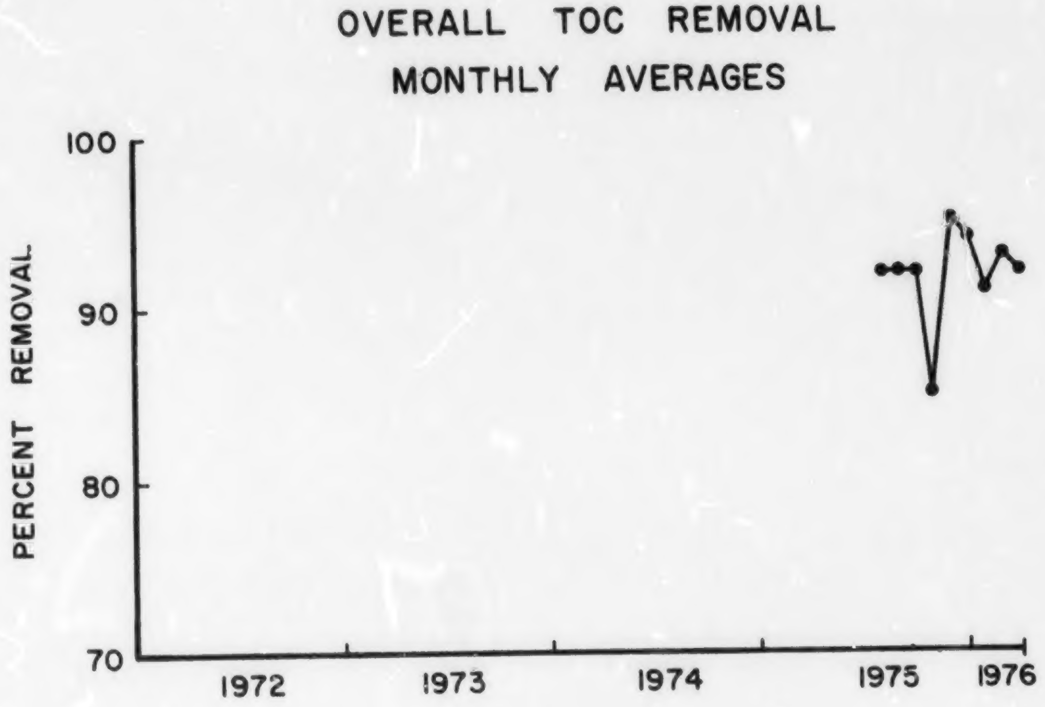


FIGURE IV-18B



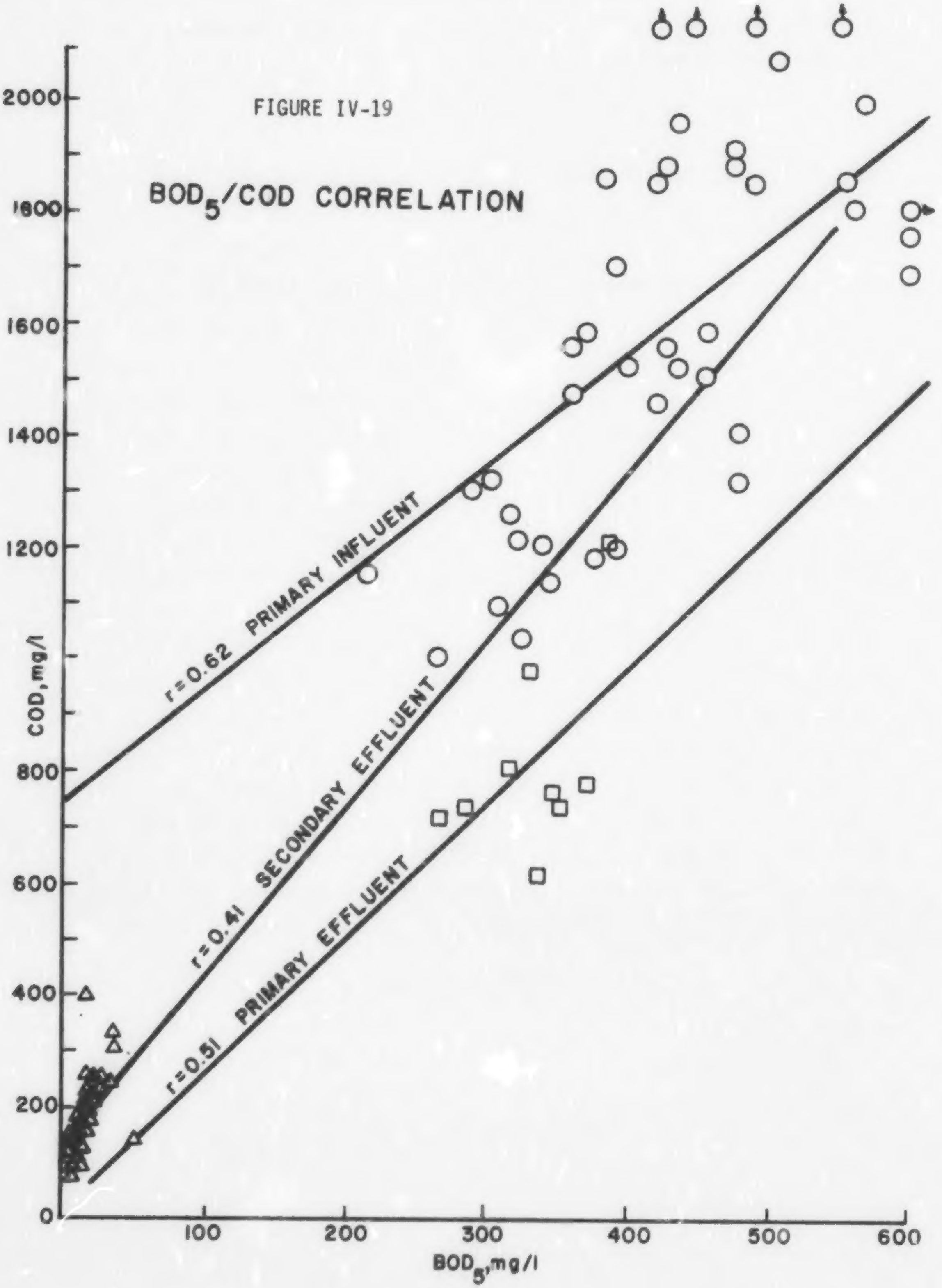


FIGURE IV-19

BOD$_5$/COD CORRELATION

It appears that there were significant amounts of inorganic solids in the plant effluent. However, for the period during coagulant addition, the relationship was:

$$TSS = 0.93\ (BOD_5) + 0.5$$
$$r = 0.83$$

Therefore, coagulant addition practices have significantly improved inorganic suspended solids removal capabilities.

DESIGN LOADINGS FOR ADVANCED WASTEWATER TREATMENT PROCESSES

The design loadings developed in this section will be used in Chapter VII to define conceptual designs for advanced wastewater treatment processes. Projected wastewater flow rates outlined in Chapter III revealed that the future discharge to the wastewater treatment plant will be approximately 205 gpm. Since the flow values used to generate that number are accurate to ± 10%, advanced wastewater treatment units should be designed to receive a minimum of 205 gpm (110%), or 226 gpm. As has been mentioned previously, it is anticipated that during winter operations, wastewater discharge will exceed reuse requirements, and it may be necessary to store reclaimed wastewater for use during the summer. It is desirable to include in the recirculation scheme provisions for treating the stored wastewater prior to reuse. The projected difference between wastewater discharge and reuse requirements is 54.3 gpm, thus, the design flow would be (205 gpm + 54.3 gpm) 110% or 285 gpm. The 12 month period from June, 1975, to May, 1976, is chosen as the period of record for determining organic and solids loads, since most of the existing facility improvements (operational and physical) had been implemented by May, 1975. The design loadings are summarized in Table IV-4.

TABLE IV-4

DESIGN LOADINGS FOR ADVANCED WASTEWATER TREATMENT

Design Flow = 285 gpm
= 410,400 gal/day

TOC Concentration = 34 mg/l
(Range: 22 mg/l - 75 mg/l)

TOC Loading = 116 lb/day

TSS Concentration = 9 mg/l
(Range: 5 mg/l - 60 mg/l)

TSS Loading = 31 lb/day

CHAPTER V

WASTEWATER TREATABILITY STUDIES

Bench and pilot scale treatability studies afford an economical means of evaluating the performance of wastewater treatment processes by observing system responses under various environmental and physical conditions. The necessity for such studies is underscored when dealing with industrial wastewaters, where "handbook" design inevitably results in additional costs either through overdesign or failure of the facility to perform as required.

Several approaches may be employed in evaluation of the individual processes which comprise a total waste treatment system. It should be recognized, however, that regardless of the approach taken, the ultimate accuracy of the information desired depends on several conditions. They include:

1. The characteristics of the wastewater used in the treatability tests are representative of those anticipated in the full scale operation;
2. the physical nature of the bench or pilot scale process is similar to that of the full scale units;
3. independent and dependent variables are considered; and,
4. environmental parameters affecting process efficiency are defined.

Observing these and other guidelines, bench and pilot scale simulation techniques can provide process information with respect to process applicability, establishment of predictor relationships, and approximation of process capacity. Although the information gained during these studies must be applied in a judicious manner, a treatability study which is properly programmed and carefully implemented does provide the basis of the logical development of unit process selection, design, and predictive performance.

The scope of the treatability studies included an evaluation of primary treatment processes (coagulation and dissolved air flotation) as well as tertiary treatment processes (sand filtration, activated carbon adsorption, ozonation, chlorination and ion exchange. The equipment utilized, the operational and analytical procedures followed, and the results obtained are presented in the following sections of this chapter.

COAGULATION

The purpose of primary treatment at the Anderson Plant is the removal of suspended material from the wastewater. The success of suspended solids removal in the primary treatment process determines, to a large degree, the effluent quality of the secondary biological system and in addition will effect the effluent quality of any tertiary system. At the time of the treatability studies, the primary suspended solids removal was somewhat marginal due to uncoagulated binder; accordingly, the effluent from the biological system contained varying amounts of colloidal, non-settleable solids. Therefore, a comprehensive coagulation study was undertaken to determine which chemical coagulants were most effective in removing the suspended, colloidal material. Particular attention was given to those chemicals which, when added to the wastewater, would not significantly increase the total dissolved solids level. This was important for wastewater recirculation considerations. For example, the primary treatment scheme at the Owens-Corning Fiberglas Jackson Plant includes ferric chloride and lime for coagulation and reportedly this system has performed well. However, substantial smounts (200 mg/1 +) of dissolved solids are added to the water in the form of calcium chloride. If significant quantities of dissolved solids were added to the wastewater at Anderson during primary treatment, TDS removal would be required for total water recirculation.

Procedure

The test procedure for the coagulation studies consisted of first placing 500 ml of neutralized wastewater into each of six beakers. The coagulant being tested was then added to five of the beakers in varying amounts. The wastewater was then flash mixed with a jar stirrer apparatus for three minutes at 120 rpm, slowly mixed for five minutes at 20 rpm and allowed to settle for 30 minutes. The supernatant was then analyzed for COD and turbidity.

Jar Test Data Analysis

The coagulation mechanism is heavily pH dependent as indicated in Figure V-1, which shows the effect of pH on colloidal stability, as measured by supernatant turbidity. No chemicals other than acid or base were added to these samples prior to coagulation. Coagulation efficiency, as measured by turbidity, is greatly impaired if the pH is less than 5.5 or greater than 7.0. For this reason all jar contents were adjusted to pH 6.7 in this study.

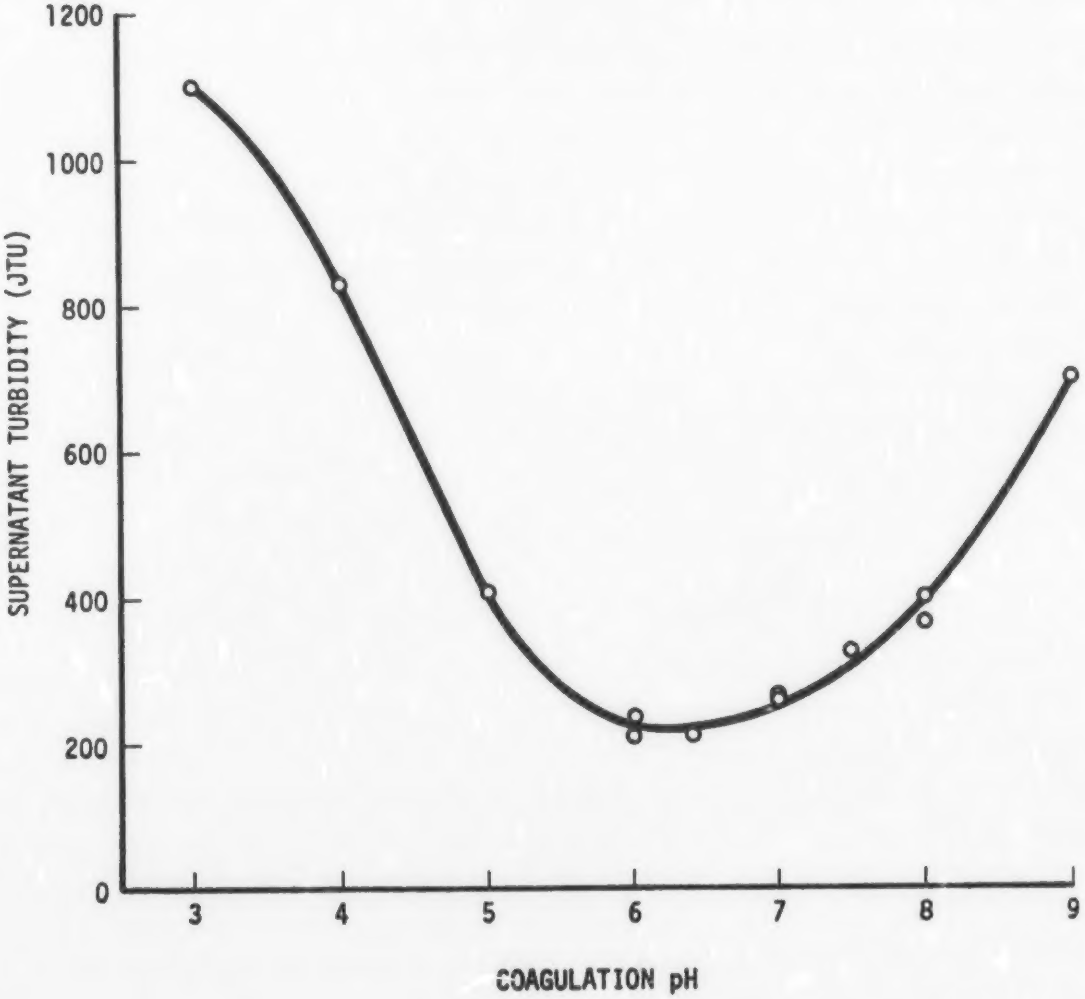
Preliminary jar tests indicated that alum was ineffective as a coagulant. Therefore, attention was directed toward ferric chloride using lime or caustic for neutralization, and synthetic polymers and clays. Tests were run using each coagulant alone and in various combinations to determine the optimum dosage of the chemicals singly or in combination. A summary of the results of these tests is presented in Figure V-2. The graph shows the COD and turbidity of the unfiltered supernatant for the optimum dose of each coagulant or combination of coagulants. Each test was performed with portions of a single wastewater sample which makes comparison of jar test results more meaningful. The results indicate that the following three coagulation schemes will produce approximately the same quality supernatant in terms of COD and turbidity.

1. 100 to 200 mg/l $FeCl_3$ alone.
2. 5 to 10 mg/l cationic polymer with 50 to 100 mg/l clay.
3. 30 to 50 mg/l $FeCl_3$ plus 5 mg/l cationic polymer and 50 mg/l clay.

However, the second coagulation scheme will minimize the addition of total dissolved solids (TDS) to the wastewater. Coagulation with ferric chloride (and caustic to neutralize to pH 6.7) substantially increases the TDS in the wastewater. For example, a sample of waste was coagulated with 250 mg/l $FeCl_3$, and the TDS approximately doubled, from 336 mg/l to 612 mg/l. The same sample was then coagulated with 10 mg/l cationic polymer and 100 mg/l clay, and the TDS actually decreased to 280 mg/l. As mentioned above, from a wastewater renovation standpoint, the polymer and clay coagulation scheme is the one of choice. However, facilities for adding ferric chloride should be installed because during upset

FIGURE V-1

EFFECT OF pH ON COLLOIDAL STABILITY*



*Simple pH adjustment through addition of acid or caustic conditions, ferric chloride may be necessary to completely coagulate batch dumps of binder.

Trial of Coagulation Scheme

In October, 1974, a full scale trial of the recommended clay polymer coagulation scheme was conducted at the Anderson Plant. The purpose of this trial was two-fold. First, it was necessary to produce a sludge typical of a clay-addition system in order to gather data for a sludge handling study reported under a separate contract. Second, it was desirable to demonstrate the effectiveness of coagulation on Anderson's variable waste. From October 10th to 28th, Nalco 8151 clay was fed at a rate of 350 lb/day and Nalco 600 polymer at 33 lb/day, which corresponded to dosages of 96 mg/l and 9.1 mg/l, respectively. Mean values for treatment plant influent, primary clarifier effluent and secondary clarifier effluent COD and suspended solids from the treatment plant log for this period are recorded in Table V-1 along with the averages of these parameters for the two weeks prior to the trial.

Trial Data Analysis

Removal efficiences for normal and trial conditions are shown in Table V-1, along with the percent improvements in the effluents due to the coagulation treatment. The clay and polymer additions effected a 64% reduction of primary effluent TSS from that during normal operations, resulting in a 30% improvement in the final effluent COD. Treatment plant COD loadings were exceptionally high during both the trial period and preceeding normal period. It should be noted that as the temporary clay feed system was subject to operational difficulties which led to uneven dosage and frequent breakdowns, better results would be expected with an adequately designed feed system. Jar tests run during the 1974 trial period with samples of one polymer produced better visual results at lower dosages than the Nalco 600, which did not perform as well as it had in 1973. It is thus advisable to stock several cationic polymers and to maintain a program of continuous jar testing to handle changes in the waste streams and new types of polymers.

FIGURE V-2

COAGULATION STUDY RESULTS USING

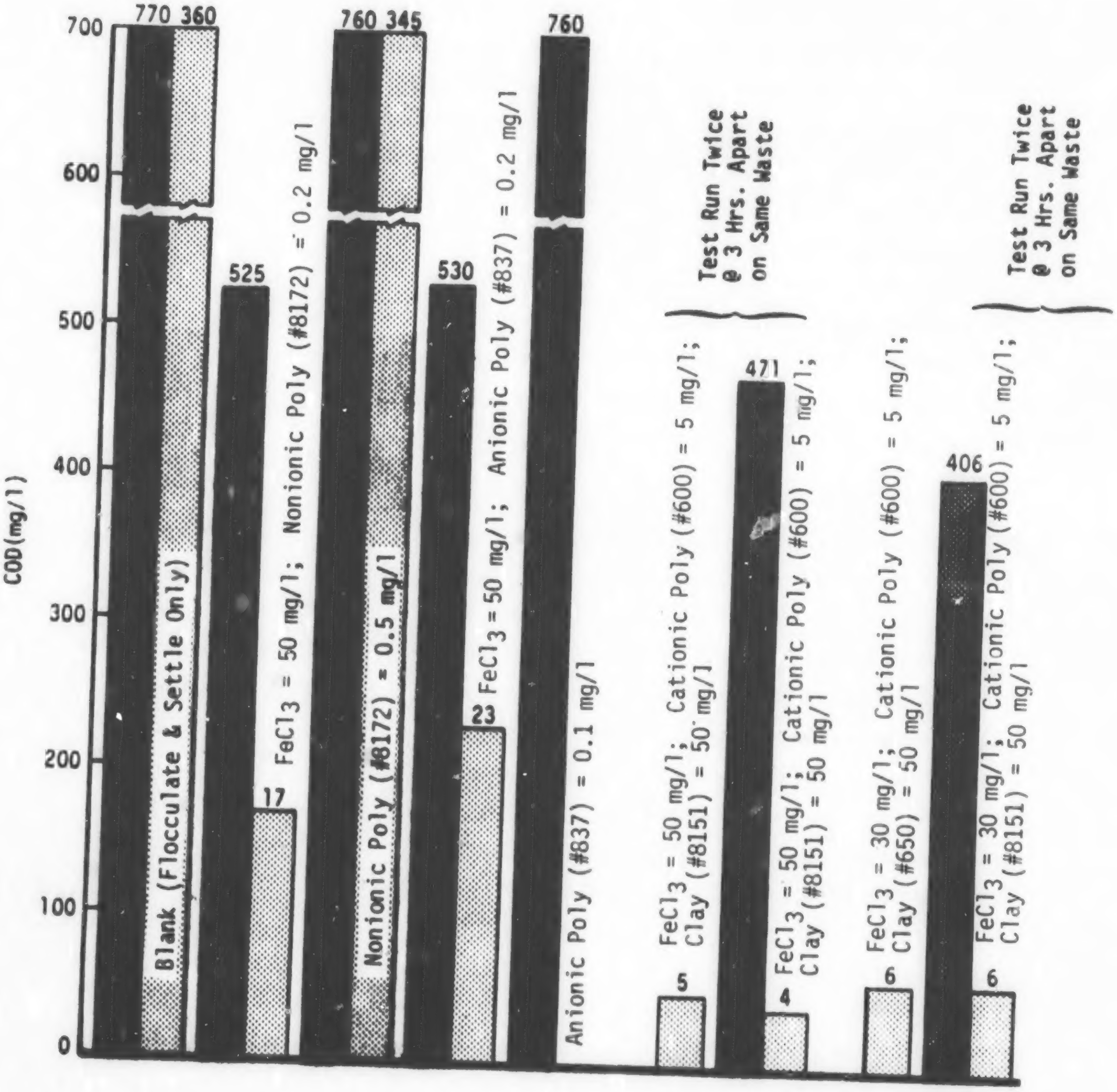


FIGURE V-2

OPTIMUM COAGULANT DOSAGES

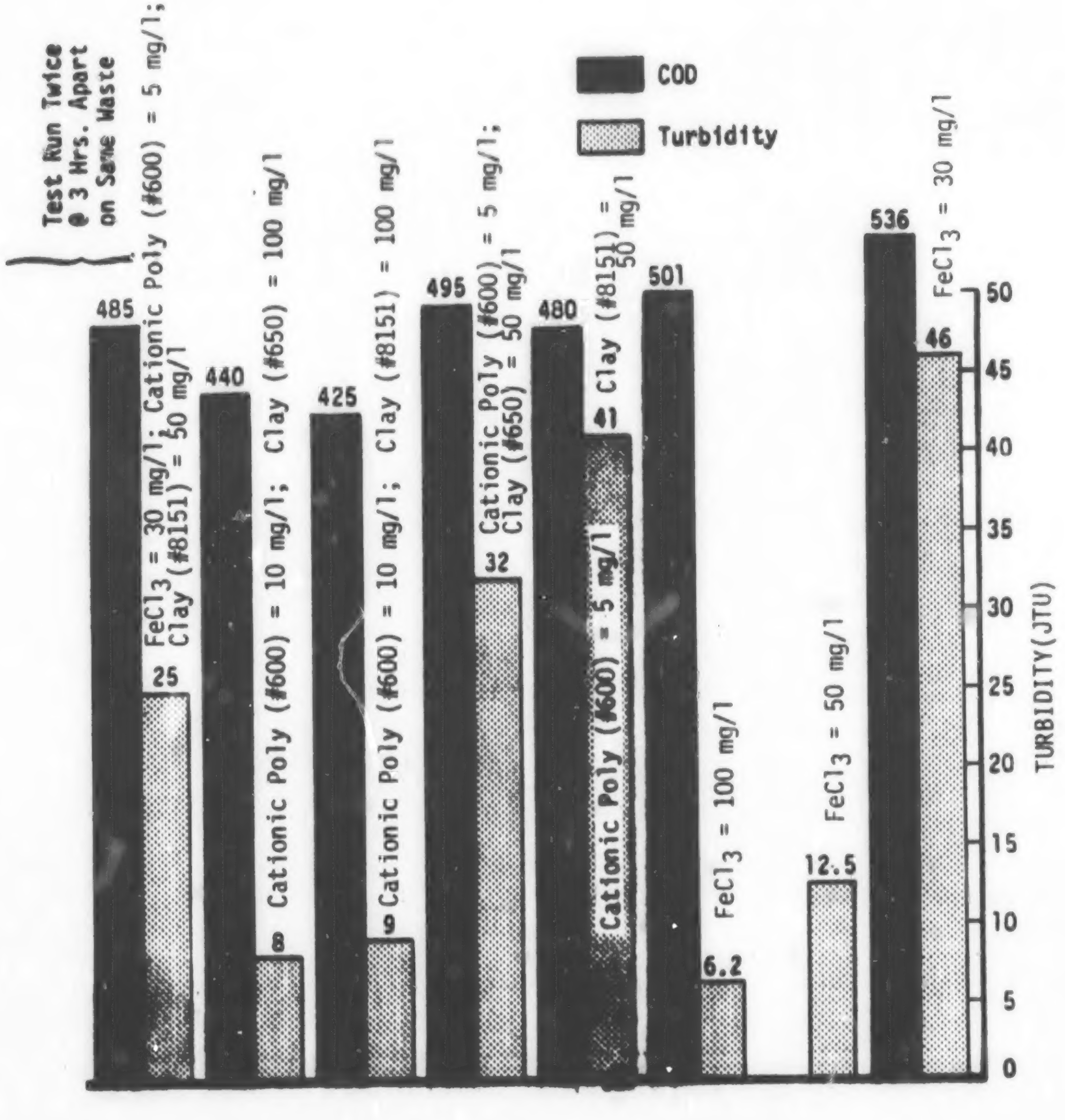


TABLE V-1

CHEMICAL FEED TRIAL RESULTS

Sample Point	Normal Conditions (9/16/74 - 10/10/74) Concentration (mg/1)	Normal Conditions (9/16/74 - 10/10/74) Removal (%)	Trial Conditions (10/11/74 - 10/28/74) Concentration (mg/1)	Trial Conditions (10/11/74 - 10/28/74) Removal (%)	Improvement (%)
Plant Influent					
COD	2064		2001		
TSS	624		638		
Turbidity	1000*		1000*		
BOD	400*		400*		
Primary Effluent					
COD	853	59	621	69	27
TSS	126	80	45	93	64
Secondary Effluent					
COD	254	88	177	91	30
TSS	14.9	98	18.5	97	-24
Turbidity	38.4	96	14.4	99	62
BOD	14.4	96	7.5	98	48

*Estimates based on 1973 data.

several cationic polymers and to maintain a program of continuous jar testing to handle changes in the waste streams and new types of polymers.

DISSOLVED AIR FLOTATION

Dissolved air flotation (DAF) is a process employed for the separation of low density suspended matter, including fibers, from a waste stream. Flotation is accomplished by pressurizing the wastewater or a recycle to 40 - 60 psig in the presence of excess air. The air, dissolved in the liquid at these increased pressures, is released from solution in the flotation unit at atmospheric pressure as minute bubbles which become attached to the particulate matter causing it to rise to the surface where it is skimmed off.

There are two potential uses for DAF at the Anderson Plant. First, as a pretreatment step for Mat Line wastewaters to remove glass fibers and binder, both of which are particularly difficult to separate by screening. Second, as an alternative to primary sedimentation for the entire waste stream. Both possibilities were investigated during these tests.

Procedure

A quantity of clarified wastewater was placed into a pressure chamber and pressurized to 60 psig. The air-liquid mixture was shaken for approximately one minute and allowed to stand for several minutes to achieve saturation. This mixture was then released into the bottom of a 1000 ml graduated cylinder partially filled with non-clarified wastewater. The effect of several coagulants was investigated by chemical addition prior to the release of pressurized water. The solids rise rate was measured and effluent quality analyses were performed on the clarified liquid. Coagulants were added to the cylinder prior to the release of pressurized water on several occasions, to investigate the effects of chemical addition.

Data Analysis - Mat Line Wastewater

The Mat Line sample was collected at a point in front of the shaker screens and contained substantial amounts of fibrous material. DAF tests were conducted using 50 percent, 100 percent, and 200 percent recycle without chemical addition, and at 100 percent recycle with the addition of several coagulants. The effluent qualities observed during these tests are shown in Table V-2. Two conclusions are immediately apparent:

1. DAF is a very effective means for treating the Mat Line wastewaters.
2. Coagulant addition is not necessary for effective removal of fibrous material.

Flotation curves for the DAF studies without coagulant addition are shown in Figure V-3. The solids rise velocity was quite high in all cases, and it was difficult to observe a distinct solid-liquid interface. Consequently, the rise velocity was conservatively estimated as shown by the slope of the broken lines.

The solids concentration in the float is a function of the air-solids ratio as shown in Figure V-4. The recycle flow can be calculated from Equation 1.

$$\frac{A}{S} = \frac{1.3 s_a R(fP-1)}{QS_a} \qquad (1)$$

where:

A/S = air/solids ratio, g air released/g solids applied
s_a = air saturation, cm^3/liter, (at 1 atm.)
R = pressurized recycle, liters/day
P = absolute pressure, atm.
Q = waste flow, liters/day
S_a = influent TSS, mg/l
f = fraction of saturation of air in the waste

For these tests:

s_a = 18.7 cm^3/l
P = 5.1 atm.
S_a = 4764 mg/l
f = 1.0 (assumed)

The float solids were quite concentrated (almost 8 percent) at air/solids ratios above .02 g air/g solids. Below this ratio the solids concentration decreased rapidly.

TABLE V-2

EFFLUENT QUALITY FOR MAT LINE

DAF STUDIES

			EFFLUENT	
TEST NO.	SAMPLE	RECYCLE	COD	TSS
1	Raw Waste (Control)	---	1315 mg/l	4764 mg/l
2	No Coagulants	50%	60	328
3	No Coagulants	100%	40	49
4	No Coagulants	100%	99	104
5	No Coagulants	200%	50	102
6	1 mg/l Polymer #607 10 mg/l Clay #650	100%	70	69
7	5 mg/l Polymer #607 50 mg/l Clay #650	100%	129	146
8	10 mg/l Polymer #607 100 mg/l Clay #650	100%	80	99
9	50 mg/l $FeCl_3$ NaOH to pH 7	100%	179	158
10	100 mg/l $FeCl_3$ NaOH to pH 7	100%	139	168
11	200 mg/l $FeCl_3$ NaOH to pH 7	100%	259	188

FIGURE V-3

MATLINE DAF

INTERFACE HEIGHT VS. RISE TIME

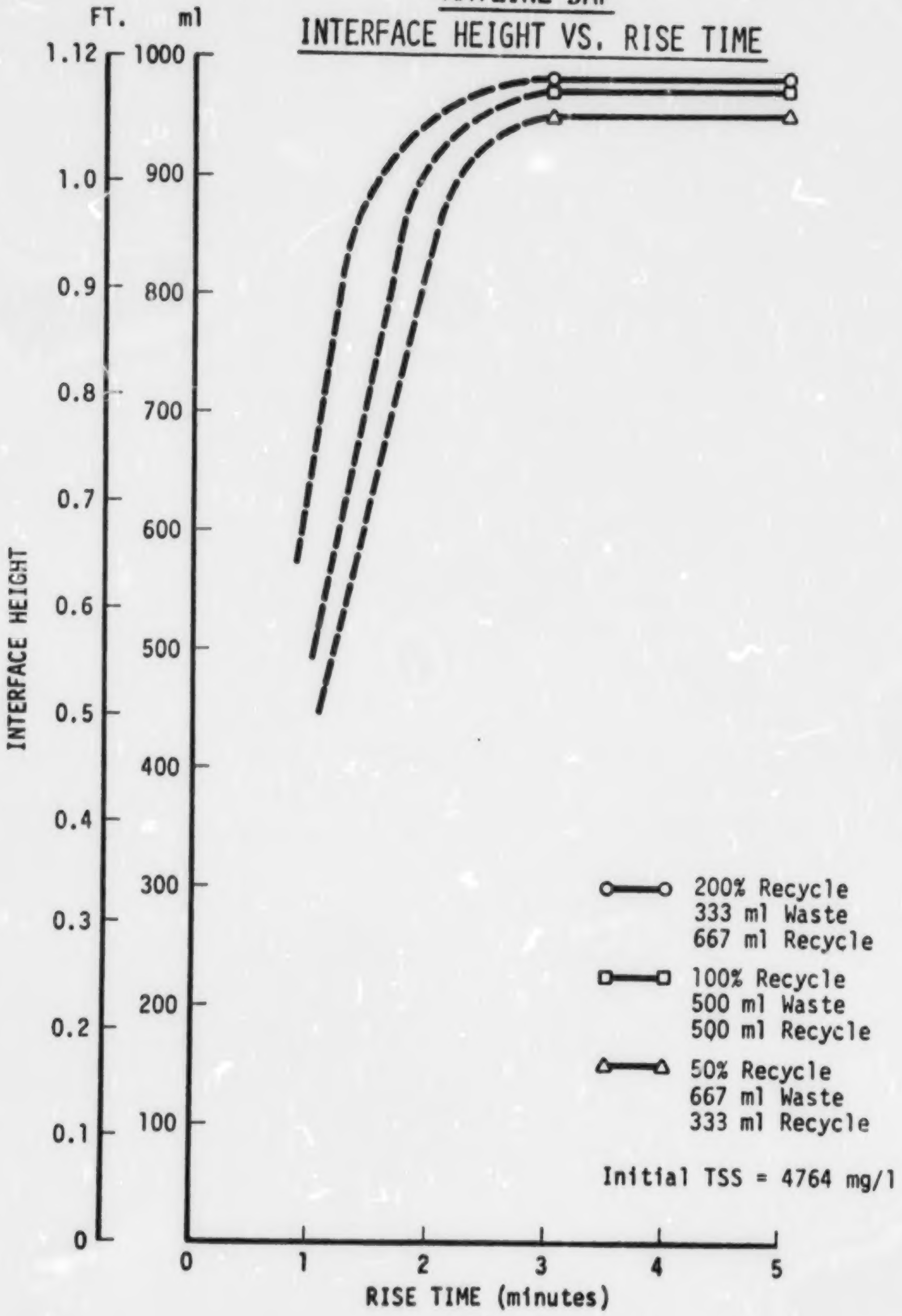
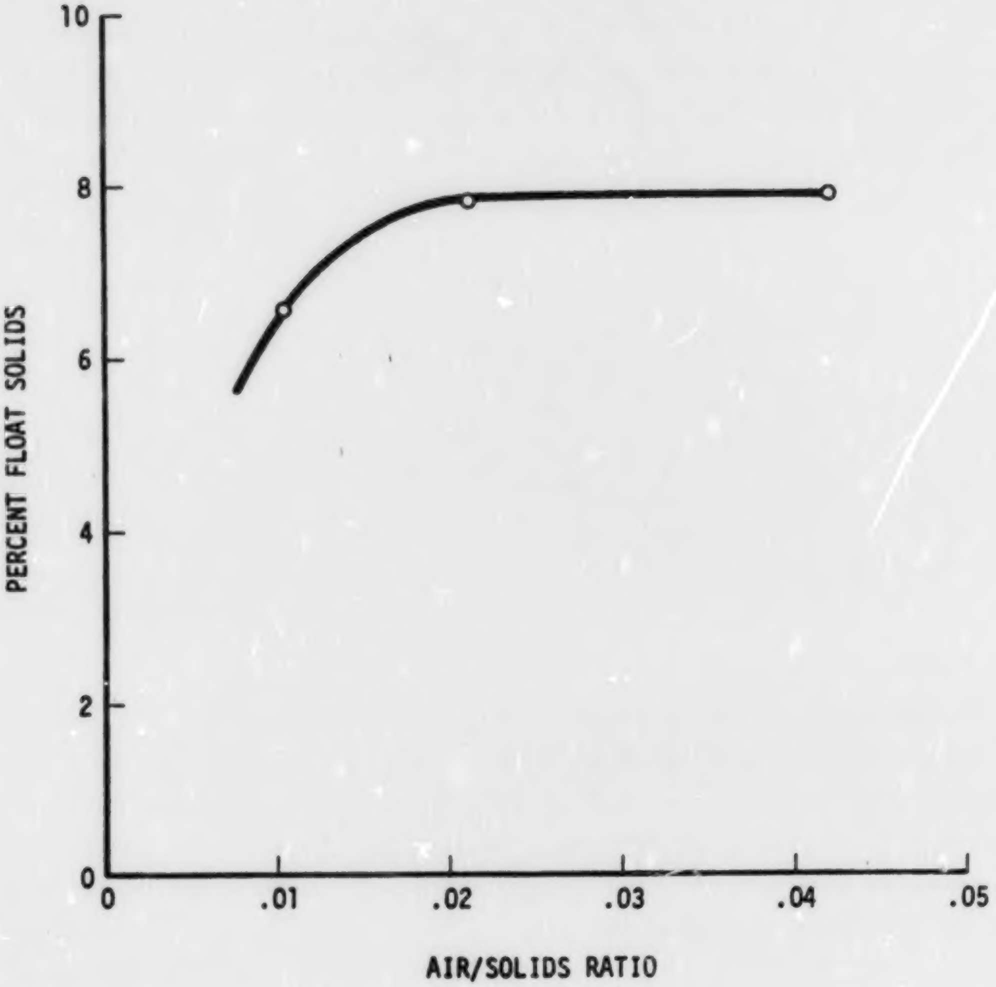


FIGURE V-4

MAT LINE DAF
PERCENT FLOAT SOLIDS VS. AIR/SOLIDS RATIO



On the basis of these test results, DAF pretreatment of the mat line wastewater will remove both glass fiber and excess binder. Design overflow rates of approximately 2 - 3 gpm/ft^2 at a recycle ratio of 100 percent at 60 psig should reduce the suspended solids and COD of the mat line wastewaters by more than 90 percent.

Data Analysis - Raw Wastewater

Dissolved air flotation was also investigated as an alternative to primary sedimentation for the entire Anderson waste stream. Two portions of a sample were treated with identical amounts of coagulants; one was settled and the other was floated with dissolved air at 100 percent recycle. The procedure was repeated for six different chemical addition schemes. Comparisons of the effluent qualities are shown in Figures V-5 and V-6. In general, DAF was more efficient than was sedimentation with respect to COD removal, while sedimentation was slightly more efficient than DAF with respect to suspended solids removal, as summarized in Table V-3.

TABLE V-3

COMPARISON OF DAF AND SEDIMENTATION

	TSS REMOVAL EFFICIENCY	COD REMOVAL EFFICIENCY
DAF	77 - 95%	69 - 80%
Sedimentation	82 - 96%	59 - 71%

Summary

Dissolved air flotation is a very effective means for removing the troublesome glass fibers from the Mat Line waste stream, and the present shaker screens should be replaced with a DAF unit. However, the differences in performance between DAF and gravity sedimentation for the entire wastewater flow are not large enough to justify replacement of the primary clarifiers with a DAF unit.

FIGURE V-5

DAF VS. SEDIMENTATION TSS REMOVAL EFFICIENCY FOR COMBINED ANDERSON WASTEWATER

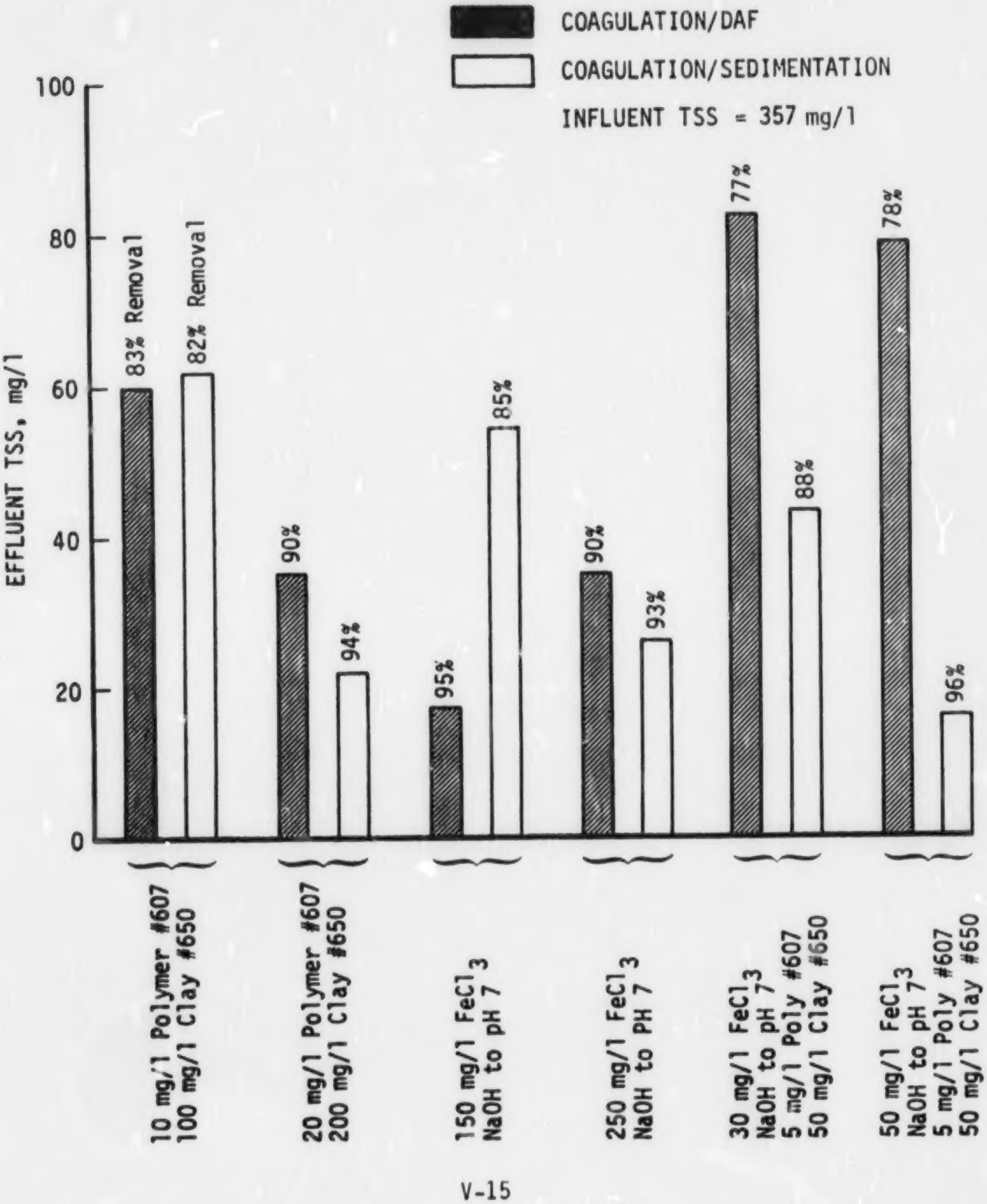
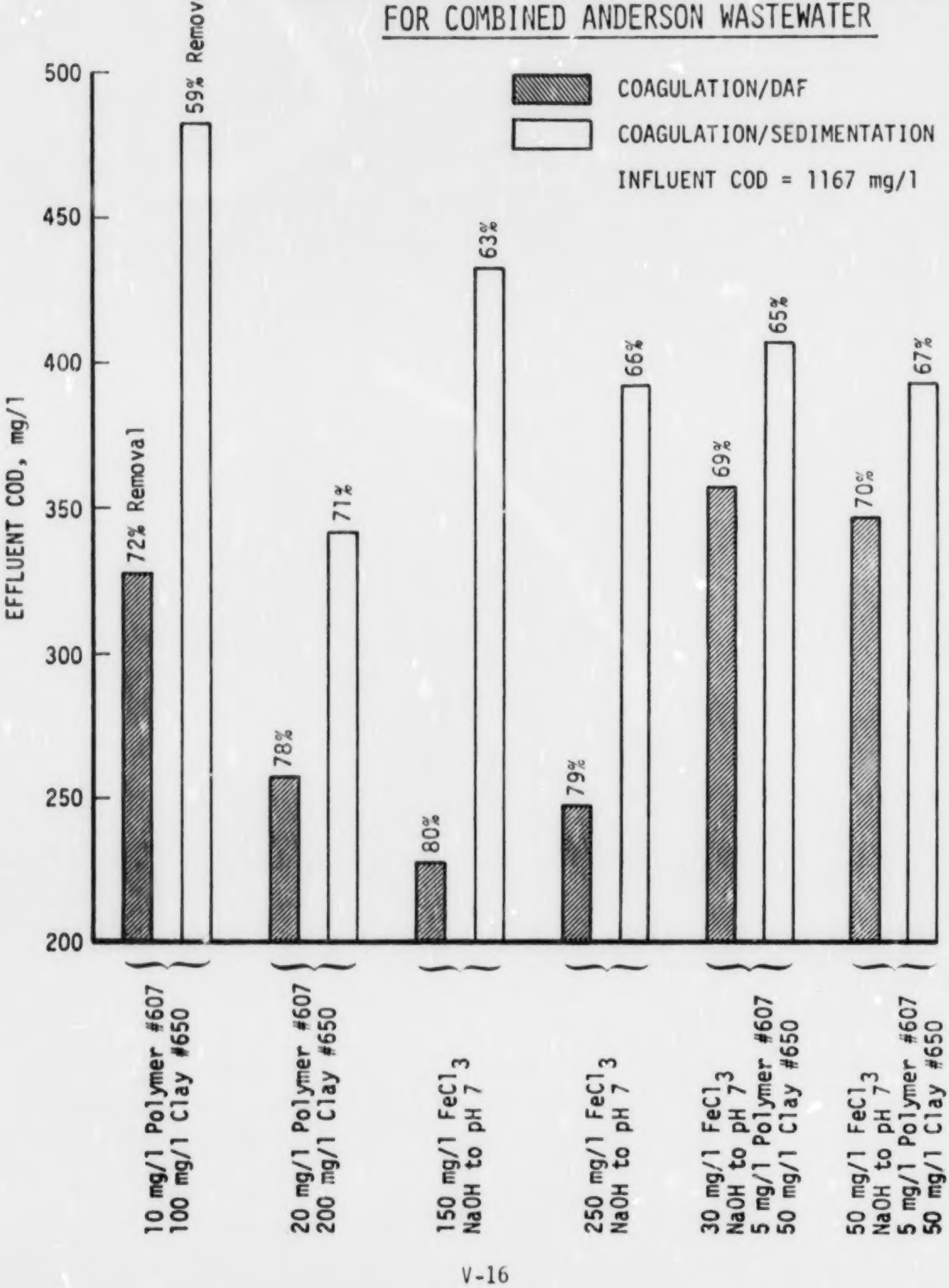


FIGURE V-6

DAF VS. SEDIMENTATION COD REMOVAL EFFICIENCY FOR COMBINED ANDERSON WASTEWATER



SAND FILTRATION

Sand filtration treatability studies were performed to evaluate tertiary filtration for removal of suspended material from the biological system effluent. Filtration will be required prior to reuse of biologically treated wastewater since most effluents of this type contain residual suspended solids.

The sand filtration pilot unit utilized for this test series was an upflow type filter as shown in Figure V-7. The filter media, graded from bottom to top, consisted of 2.5 ft^3 of 1.25 to 1.50 inches gravel, 6 ft^3 of 3/8th to 5/8ths inch gravel, 7 ft^3 of 2.0 to 3.0 mm sand, and 40 ft^3 of 0.5 to 1.0 mm sand. The filter was piped to receive biologically treated effluent from the final clarifiers. Wastewater was pumped through the filter on an upflow basis with the variable speed pump (0 - 100 gpm) supplied with the unit. This feed pump also acted as the backwash pump and was utilized in conjunction with an air blower during the backwash cycle. Samples of the filter influent and effluent were collected on a grab basis throughout each filter run.

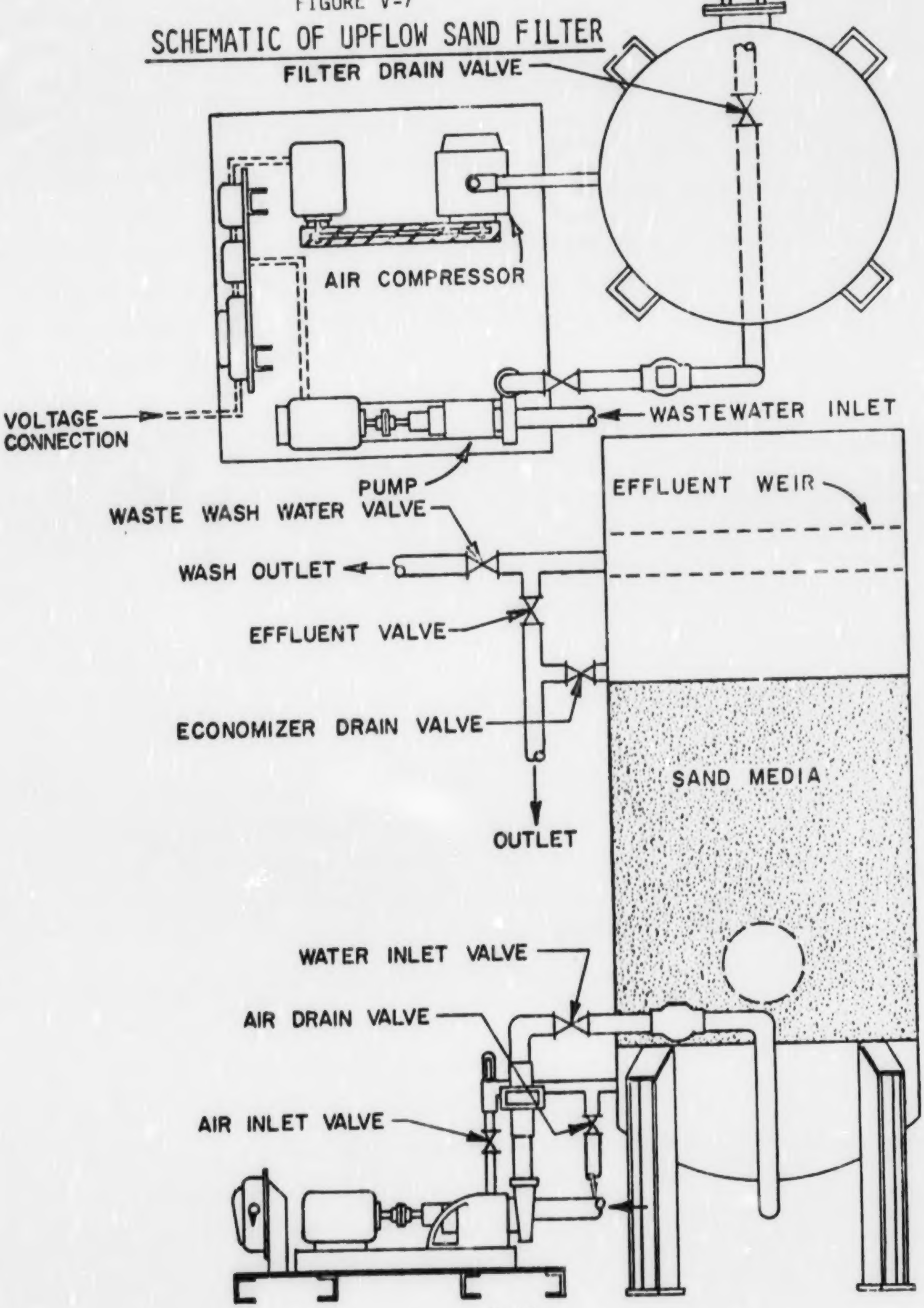
Procedures

The operational procedures used for each filter run are listed below:

1. The filter was backwashed prior to each test run. The backwash cycle included bumping the filter with 30 cfm of air for three to four minutes. The 100 gpm backwash rate was then continued for an additional 6 to 10 minutes until a clear effluent was produced.
2. The filter bed was "tightened" by draining the filter through the bottom drain valve to a water level just above the sand level.
3. The filtration cycle was initiated controlling the hydraulic flow rate manually with a valve.
4. Turbidity tests were performed on grab samples of the effluent throughout the filter run. The breakpoint was established when the turbidity reached a pre-defined level.

FIGURE V-7

SCHEMATIC OF UPFLOW SAND FILTER



Pilot sand filtration tests were performed at surface loadings ranging from 3.4 to 6.5 gpm/ft^2.

Data Analysis

A summary of the results from the pilot filtration tests is presented in Table V-4. The data from a typical filter run are presented in Figure V-8. In general, the effluent turbidity and suspended solids remained reasonably constant throughout the filter run until the actual breakthrough occurred as indicated in Figure V-8. As expected, the pressure differential across the bed gradually increased during the test run until breakthrough.

Removal efficiency through the sand filter as shown in Table V-4 is somewhat disappointing in that relatively high concentrations (above 20 mg/l) of suspended solids were present in the effluent. Sand filtration in similar applications is capable of producing effluent TSS well below 10 mg/l irrespective of the influent TSS concentration. However, considerable amounts of colloidal material were present in the final clarifier effluent and this material was not removed to any large degree by sand filtration. This underscores the necessity of adequate primary treatment for the removal of colloidal binder material present in the wastewater. Problems with the operation of the sand filter were encountered during Test No. 2 resulting from air being injected into the bottom of the filter bed. This problem was corrected during later filter runs.

ACTIVATED CARBON ADSORPTION

The feasibility of activated carbon treatment of the Anderson Plant wastewaters was evaluated from the standpoint not only of complete physical-chemical treatment but also tertiary treatment (following the activated sludge process). This section will discuss the procedures and results of bench scale adsorption isotherms and adsorption column studies.

Physical-Chemical Treatment

In the physical-chemical approach as applied to the Anderson Plant, the envisioned treatment train would consist of neutralization, coagulation, sedimentation, filtration, and granular activated carbon adsorption,

TABLE V-4

SUMMARY OF SAND FILTRATION PILOT UNIT TESTS

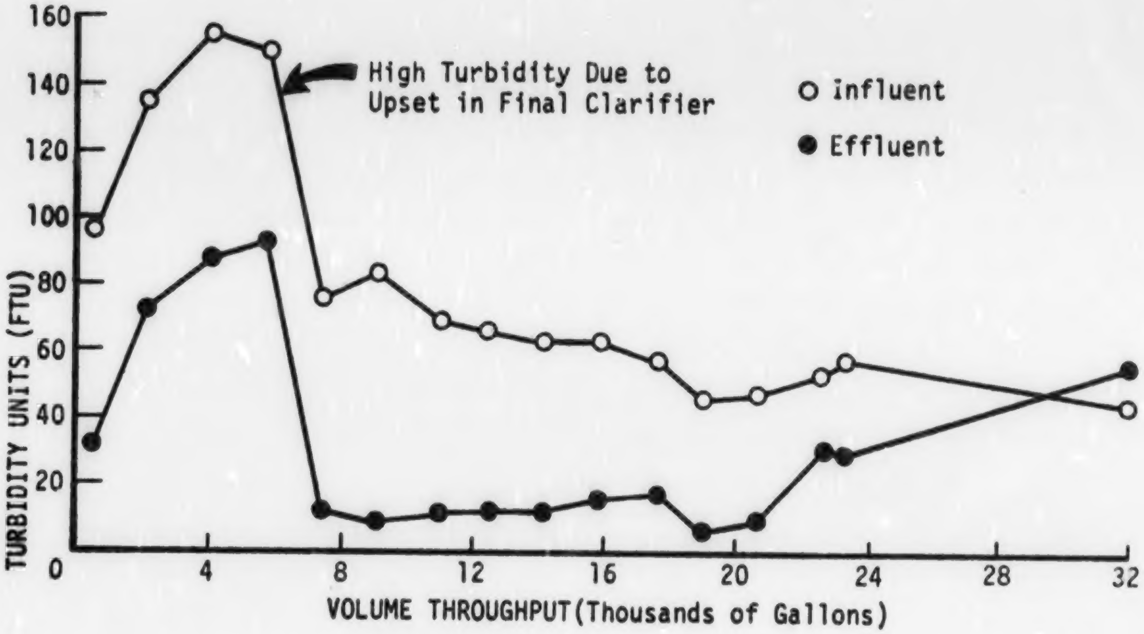
TEST NO.	DATE	FEED RATE (gpm/ft^2)	TOTAL SUSPENDED SOLIDS*		TURBIDITY*		SOLIDS STORAGE AT BREAKTHROUGH ($\frac{\text{lb TSS}}{\text{ft}^3\text{ media}}$)	TOTAL VOLUME FILTERED (gal)	NOTES
			INF. (mg/l)	EFF. (mg/l)	INF. (JTU)	EFF. (JTU)			
1	12/17/73	4.0	57	22	85	12	0.11	21,000	Initial high turbidity in influent due to interruption in polymer feed to secondary clarifiers
2	1/11/74	3.4	12	6	8	4	0.01	17,700	Test terminated due to air in filter
3	1/18/74	5.1	9	7	16	4	0.02	73,000	Clarifier effluent relatively free of colloidal suspensions until upset at termination of filter test
4	1/29/74	6.6	37	21	37	26	0.11	46,000	Algae in clarifier effluent due to sludge lagoon recycle through treatment plant
5	1/30/74	3.6	68	25	69	43	0.13	20,400	Colloidal suspended material in clarifier effluent

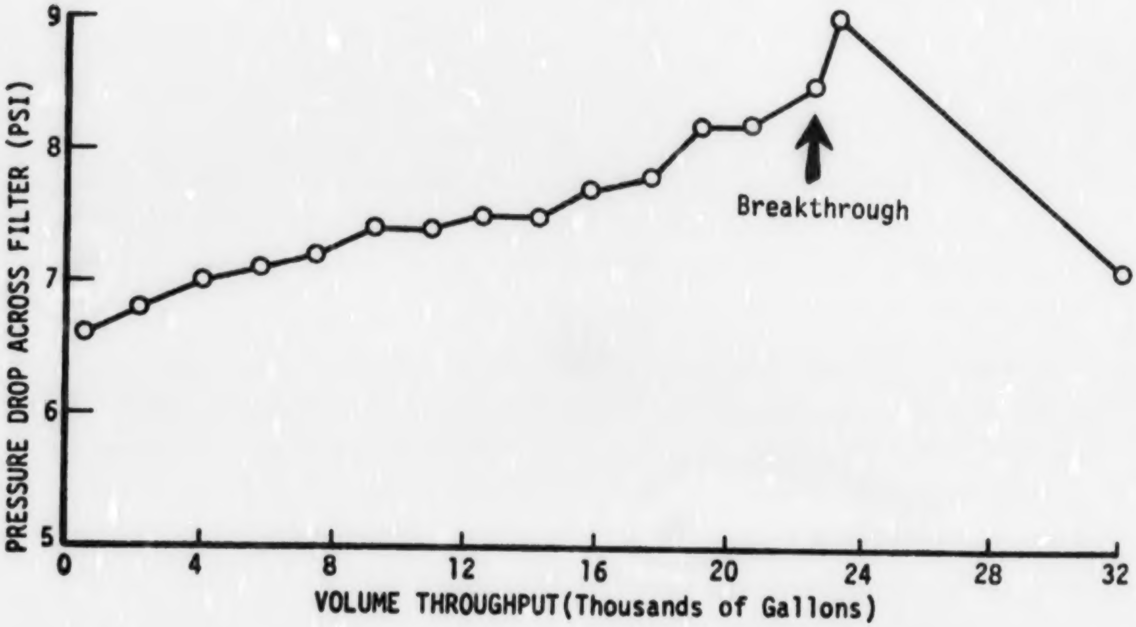
*Composited grab samples.

FIGURE V-8

UPFLOW SAND FILTER OPERATING CHARACTERISTICS

4 GPM/FT^2 - 12/17/73





thus eliminating the biological process. In cases where this scheme is effective, a number of advantages are realized. To investigate physical-chemical treatment, composited raw wastewater samples were neutralized, coagulated, and clarified. Both carbon isotherms and carbon column studies were performed on the pretreated wastewater.

Adsorption Isotherms

Adsorption isotherms were completed on samples of the pretreated Anderson Plant raw wastewater collected October 16 and 30, 1973 and on November 8, 1973. Additionally, adsorption isotherms were completed on the OCF Jackson Plant effluent as discussed below. Commercial activated carbon selected for these tests included Westvaco Aqua Nuchar A, Calgon Filtrasorb 400, and Westvaco WVL. The first two carbons are powdered material and the latter is granular. The granular carbon was crushed to a fine powder before the isotherm studies so as to increase the rate of adsorption. This was permissible because the final equilibrium adsorption values were the values of interest.

In the isotherm procedure, various weights of activated carbon were added to 500 ml quantities of filtered wastewater and the samples were mixed for 1 1/2 hours, allowing adsorption to occur. The samples were then filtered with Whatman No. 42 filter paper and the resulting COD was measured. For each sample the loading on the carbon was calculated from the formula:

$$\frac{X}{M} = \frac{(Co-C)V}{M} \qquad (2)$$

where X/M is the loading per unit weight of carbon, Co the initial concentration, C the final concentration, V the sample volume and M the carbon weight. By preparing samples with different carbon dosages, the relationship between X/M and C was developed.

Figures V-9 and V-10 present the isotherms completed on the Anderson Plant and Jackson Plant wastewaters, respectively. In both cases, the COD isotherms appear normal in that a nearly logarithmic relationship was measured between equilibrium COD and carbon loading. In fact, the Jackson isotherms are almost exact duplicates of the Anderson isotherms which indicates similar adsorption characteristics. Very high apparent

FIGURE V-9

CARBON ISOTHERMS ON ANDERSON PLANT COAGULATED RAW WASTEWATER

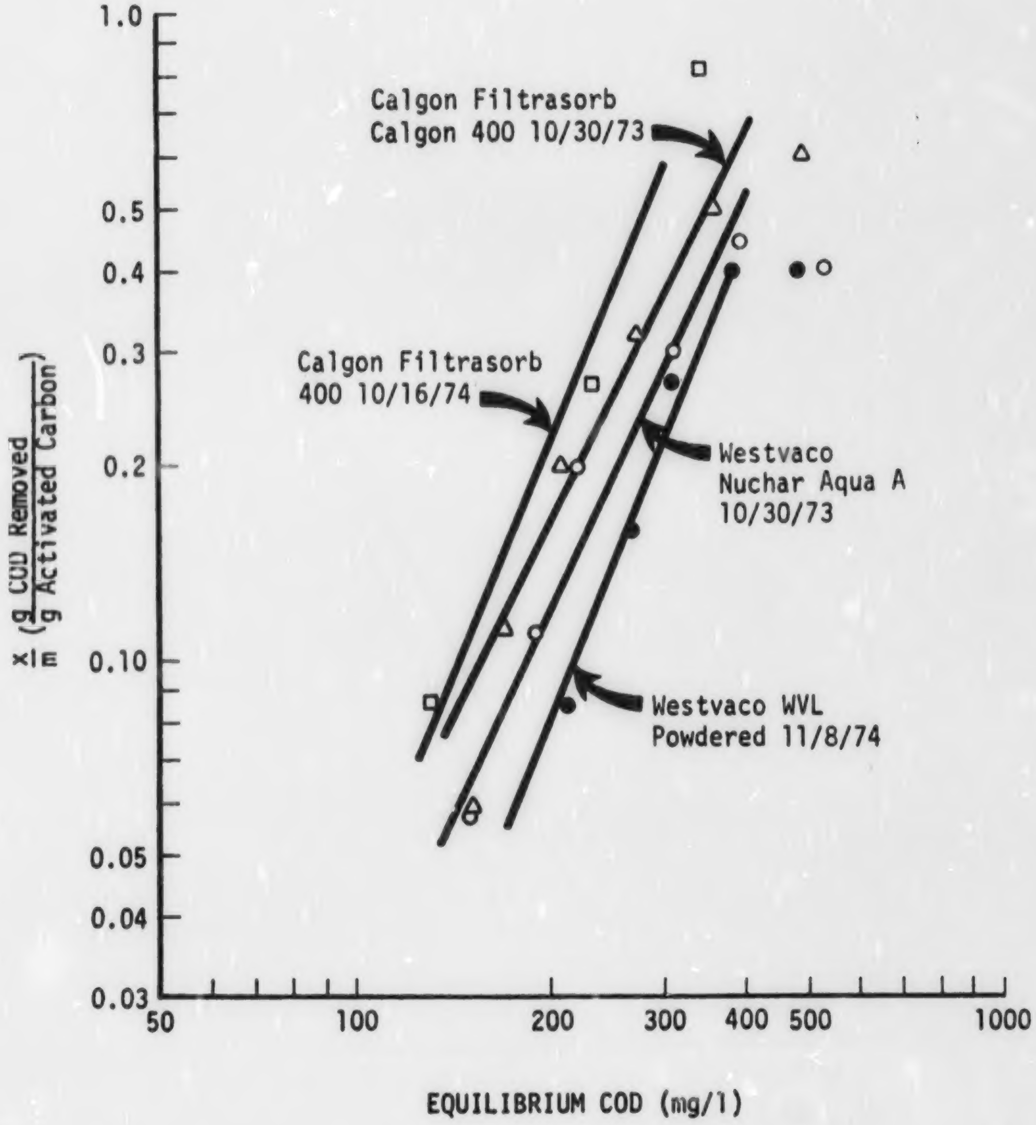
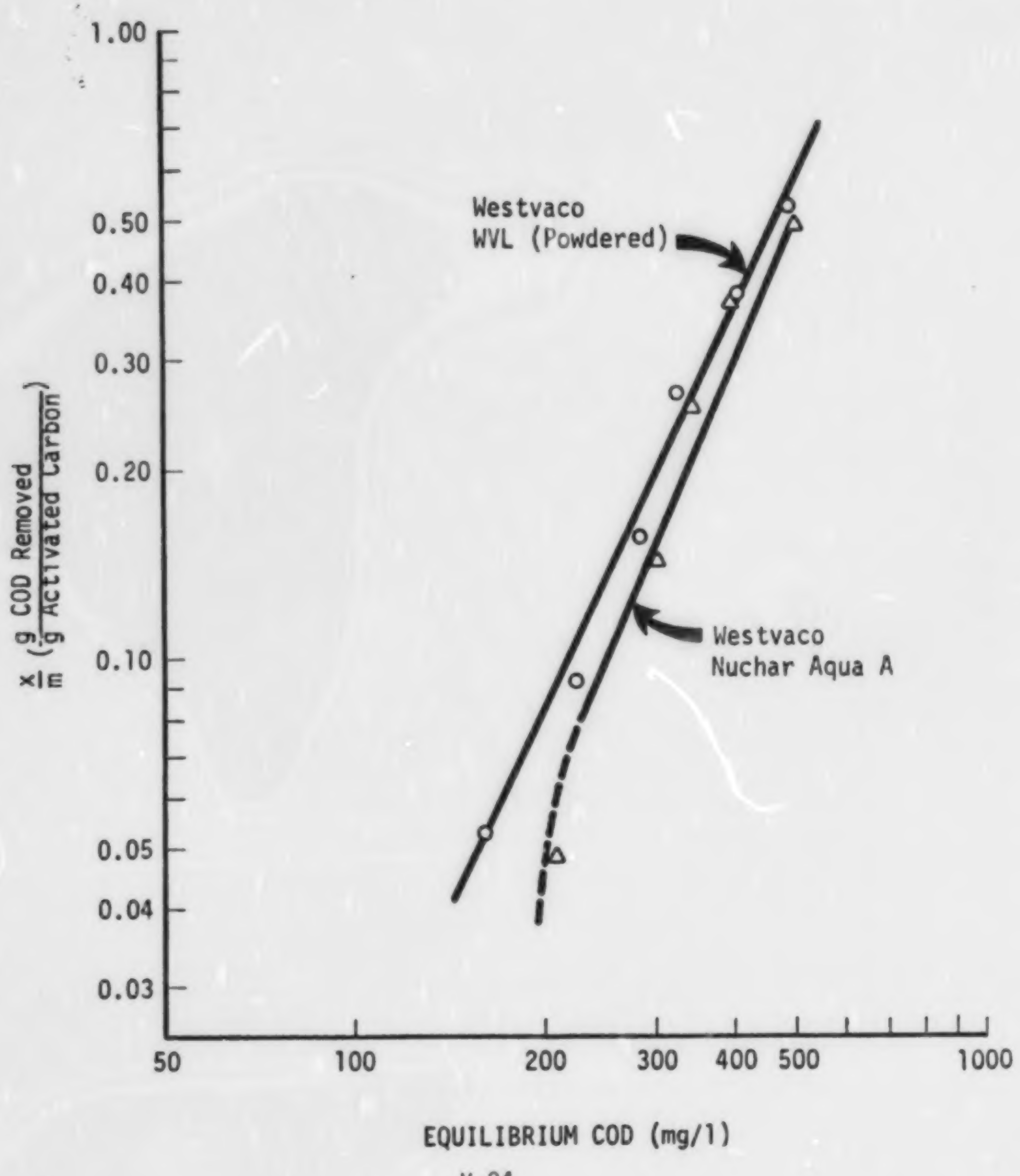


FIGURE V-10

CARBON ISOTHERMS ON JACKSON PLANT EFFLUENT NOV. 16, 1973



COD loadings were obtained for all three carbons. Residual (non-adsorbable) COD was not apparent at these dosages although at the lower COD loadings, equilibrium COD values range between 130 and 250 mg/l. Such a range compares favorably with the column studies discussed below.

Carbon Column Studies

Since the isotherms showed encouraging results, column studies were conducted to better define the effectiveness of physical-chemical treatment. Six 2.9 inch I.D. fiberglass columns each six feet in length (in series) and associated stainless steel tubing and valves were the major elements of testing equipment. Prior to beginning an experiment, each column was loaded with 5.5 pounds of activated carbon. A small variable speed pump fed the columns and a rotameter was used for flow measurement. The hydraulic loading for these tests was 4 gpm/ft^2. Columns could be backwashed at essentially any desired flow rate using the variable speed pump. Effluent from the final column was collected and stored for backwashing.

The results of the carbon column experiment in terms of COD and TOC removal are presented graphically in Figures V-11 through V-17. Figures V-11 and V-12 reflect the effluent COD levels from each of the six columns which were operated in series. As shown in Figure V-12, effluent COD levels of approximately 200 mg/l are obtainable by the physical-chemical process. Similarly, minimum effluent TOC levels are approximately 75 mg/l as shown in Figures V-15 and V-16. This relatively high discharge of organic material indicates that sizable fractions of COD and TOC are poorly adsorbed resulting in rapid columnar breakthrough. While no analyses were performed to identify the poorly adsorbed materials, materials which generally exhibit this behavior are low molecular weight, highly oxygenated or ionized organics; however, these materials are very often highly biodegradable. For example, the raw wastewater BOD_5 on one set of column samples was 162 mg/l while the effluent BOD_5 of the last carbon column was 82 mg/l, respectively. In this case, the BOD removal was only 50%.

On the basis of these results, physical-chemical treatment does not appear to be attractive because of the relatively high leakage of biodegradable organic material present in the wastewater. From a carbon

FIGURE V-11

COD VS. VOLUME THROUGHPUT - PHYSICAL-CHEMICAL TREATMENT (COLUMNS 1, 2 & 3)

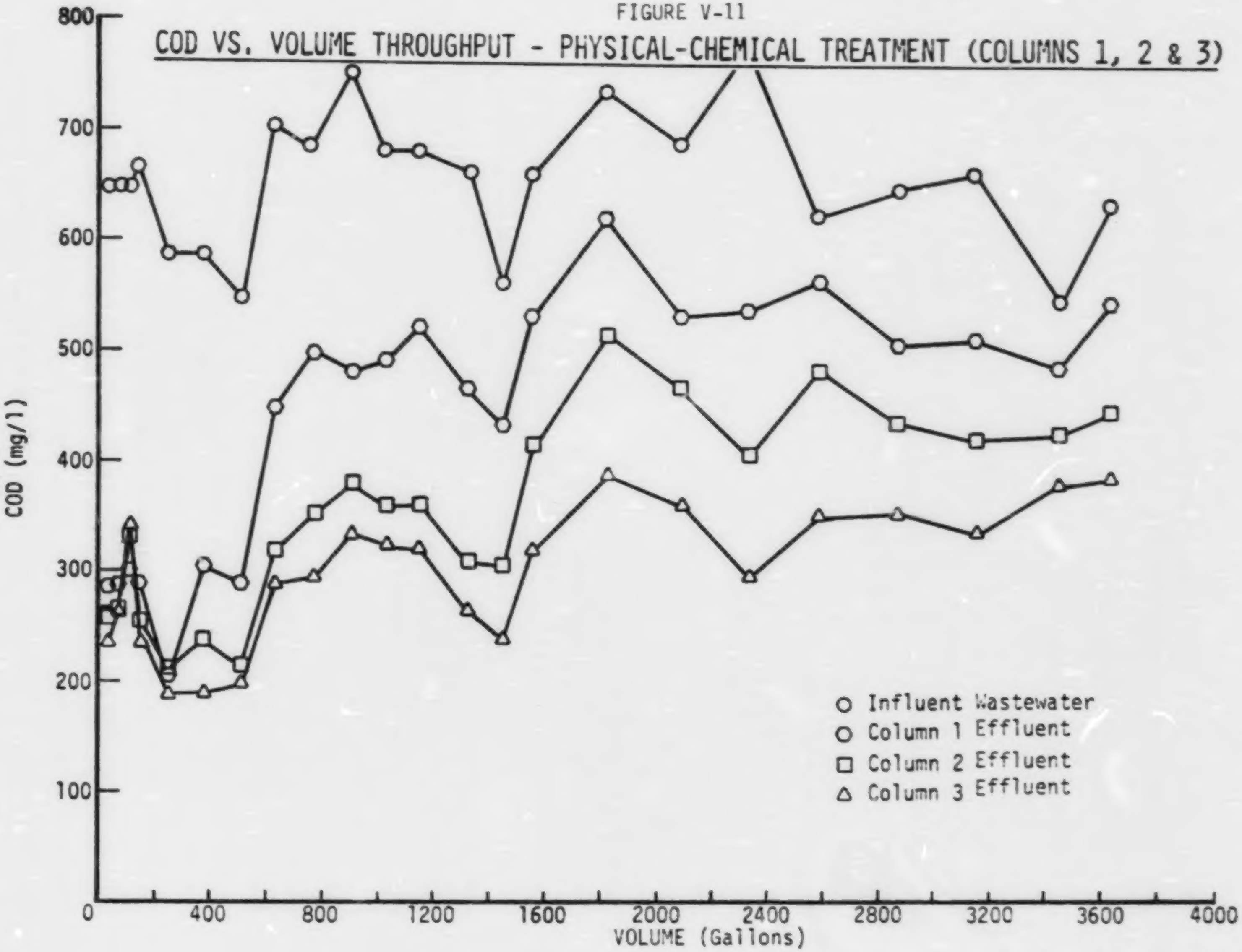


FIGURE V-12

COD VS. VOLUME THROUGHPUT - PHYSICAL-CHEMICAL TREATMENT (COLUMNS 4, 5 & 6)

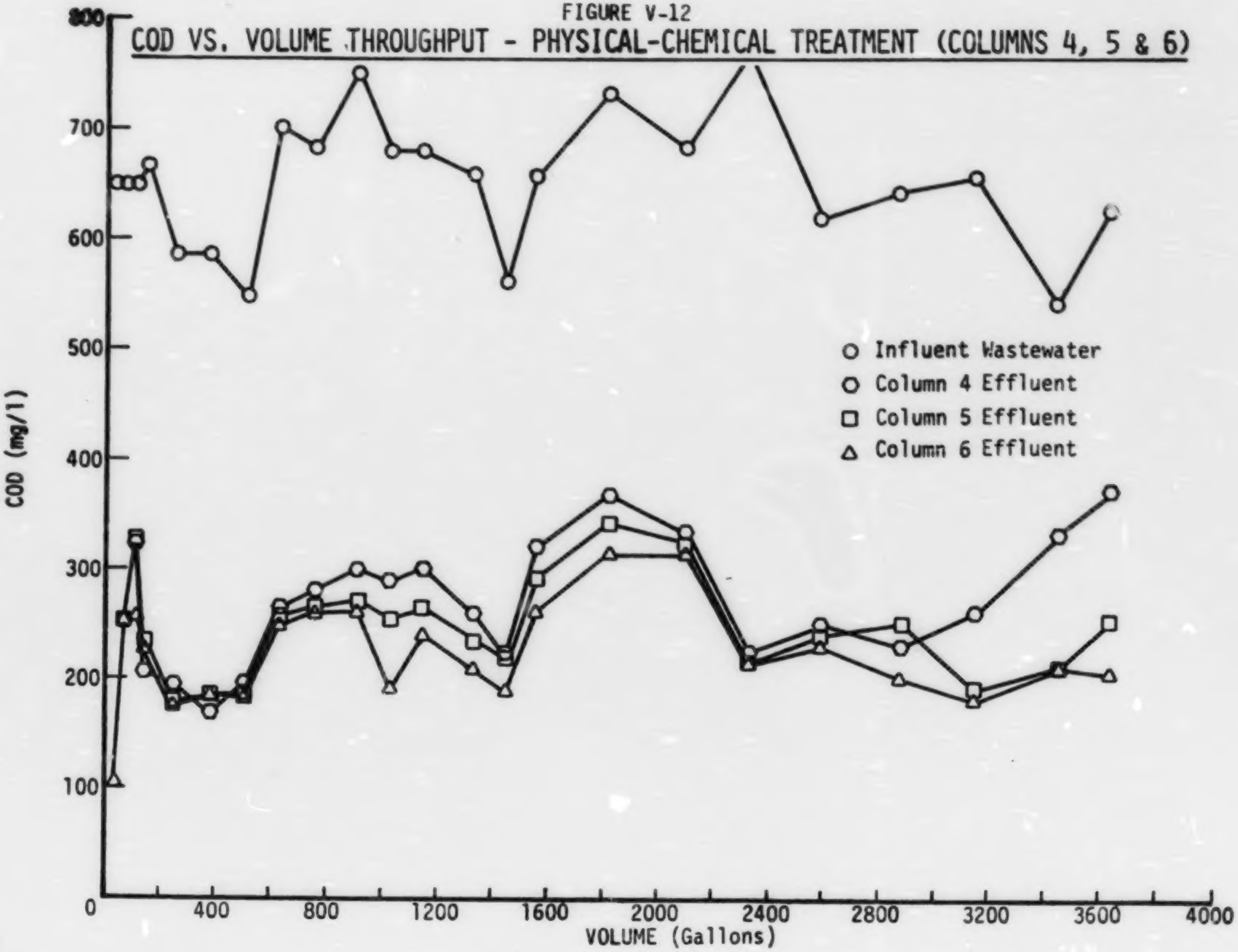


FIGURE V-13

% COD REMOVAL VS. VOLUME THROUGHPUT - PHYSICAL-CHEMICAL TREATMENT (COLUMNS 1, 2 & 3)

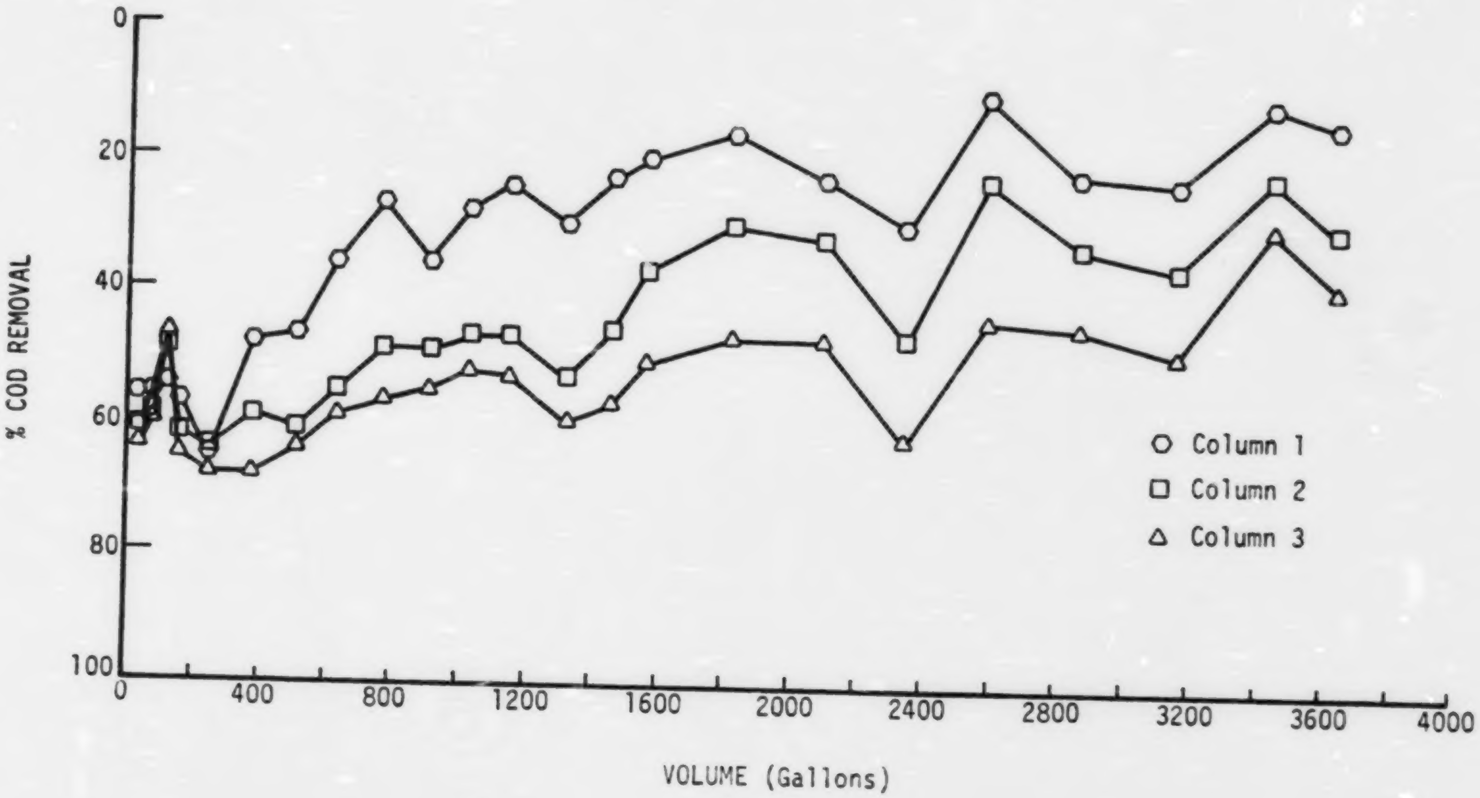


FIGURE V-14

COD LOADING VS. VOLUME THROUGHPUT - PHYSICAL-CHEMICAL TREATMENT

(COLUMN 1, 2 & 3)

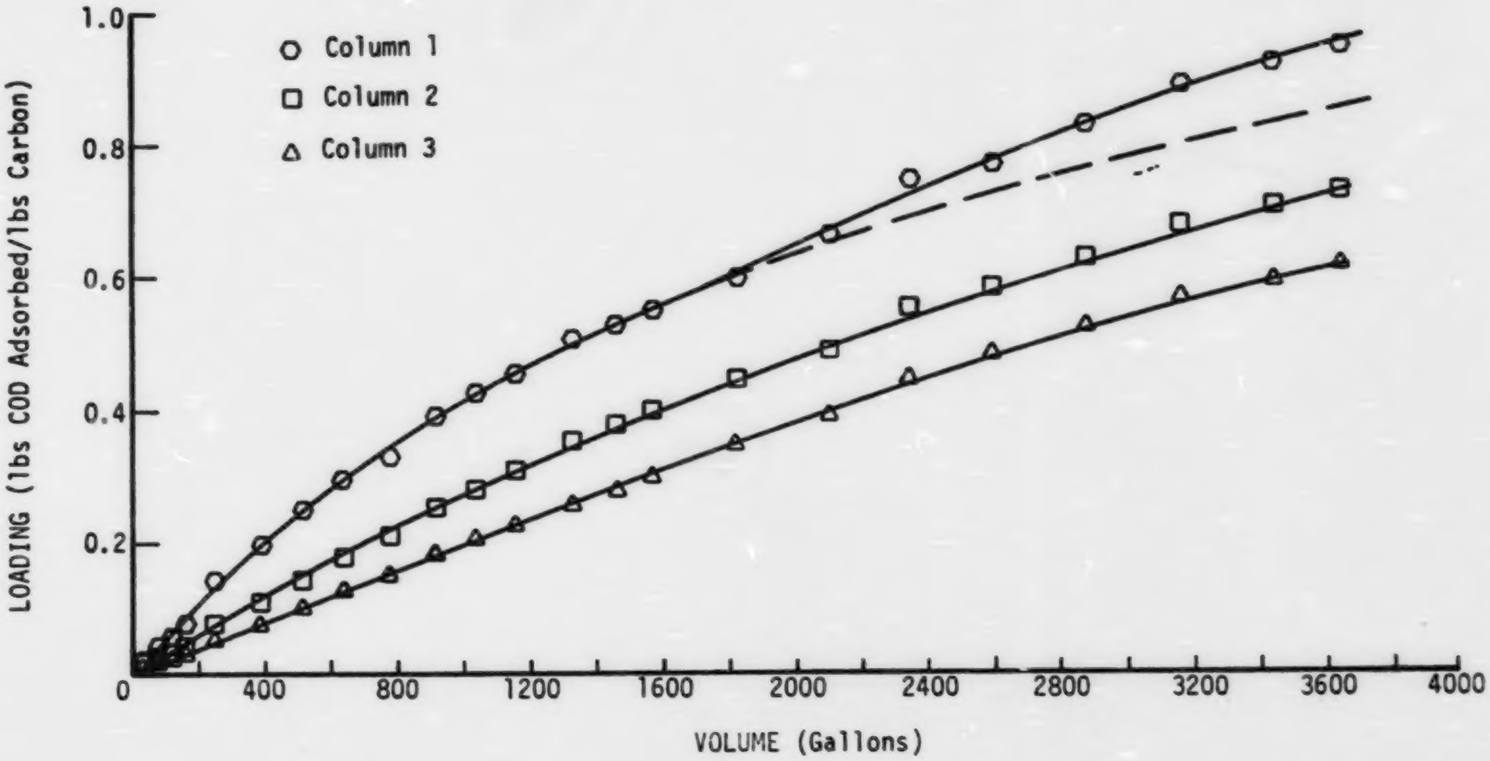


FIGURE V-15

TOC VS. VOLUME THROUGHPUT - PHYSICAL-CHEMICAL TREATMENT

(COLUMNS 1, 2 & 3)

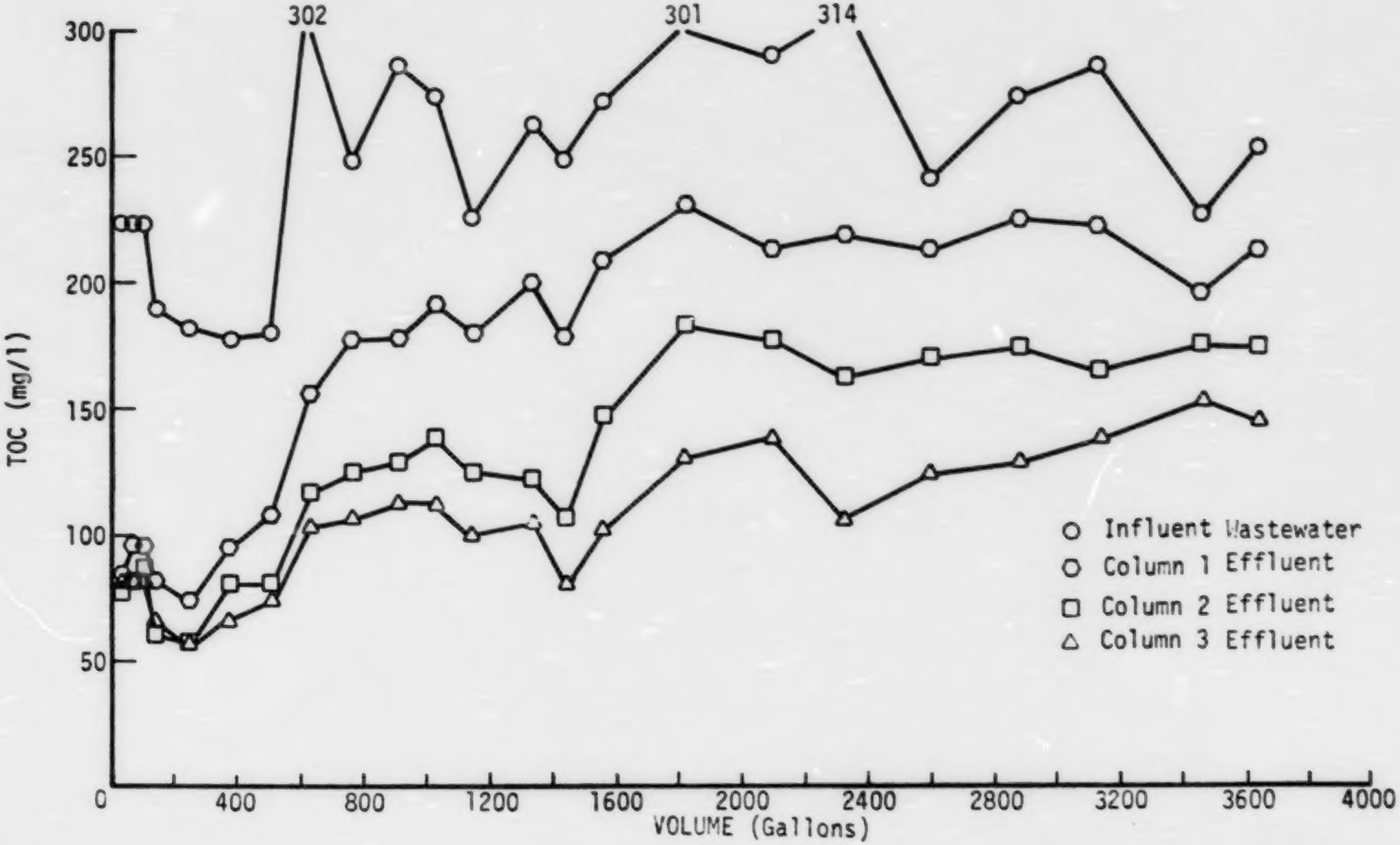


FIGURE V-16

TOC VS. VOLUME THROUGHPUT - PHYSICAL-CHEMICAL TREATMENT

(COLUMNS 4, 5 & 6)

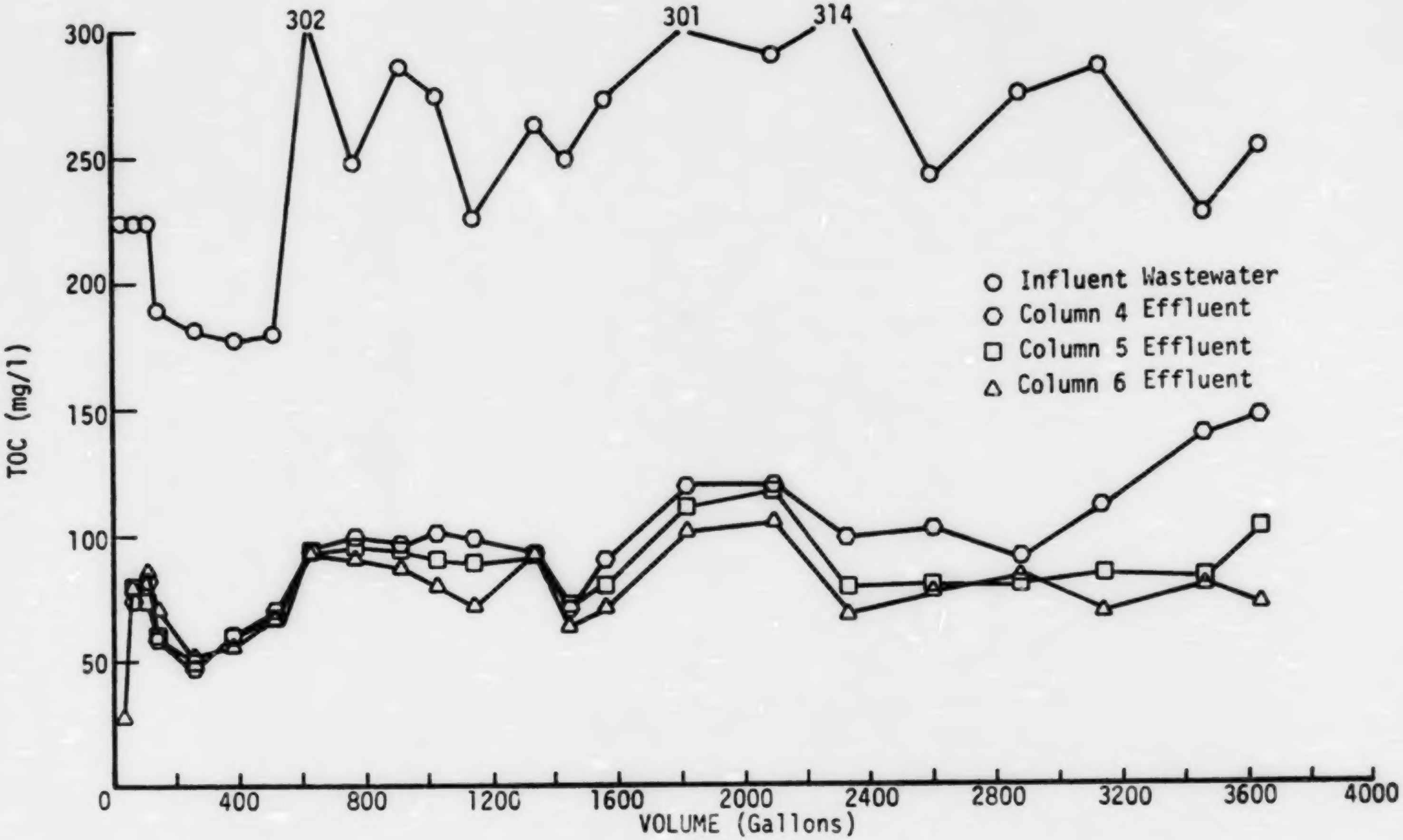
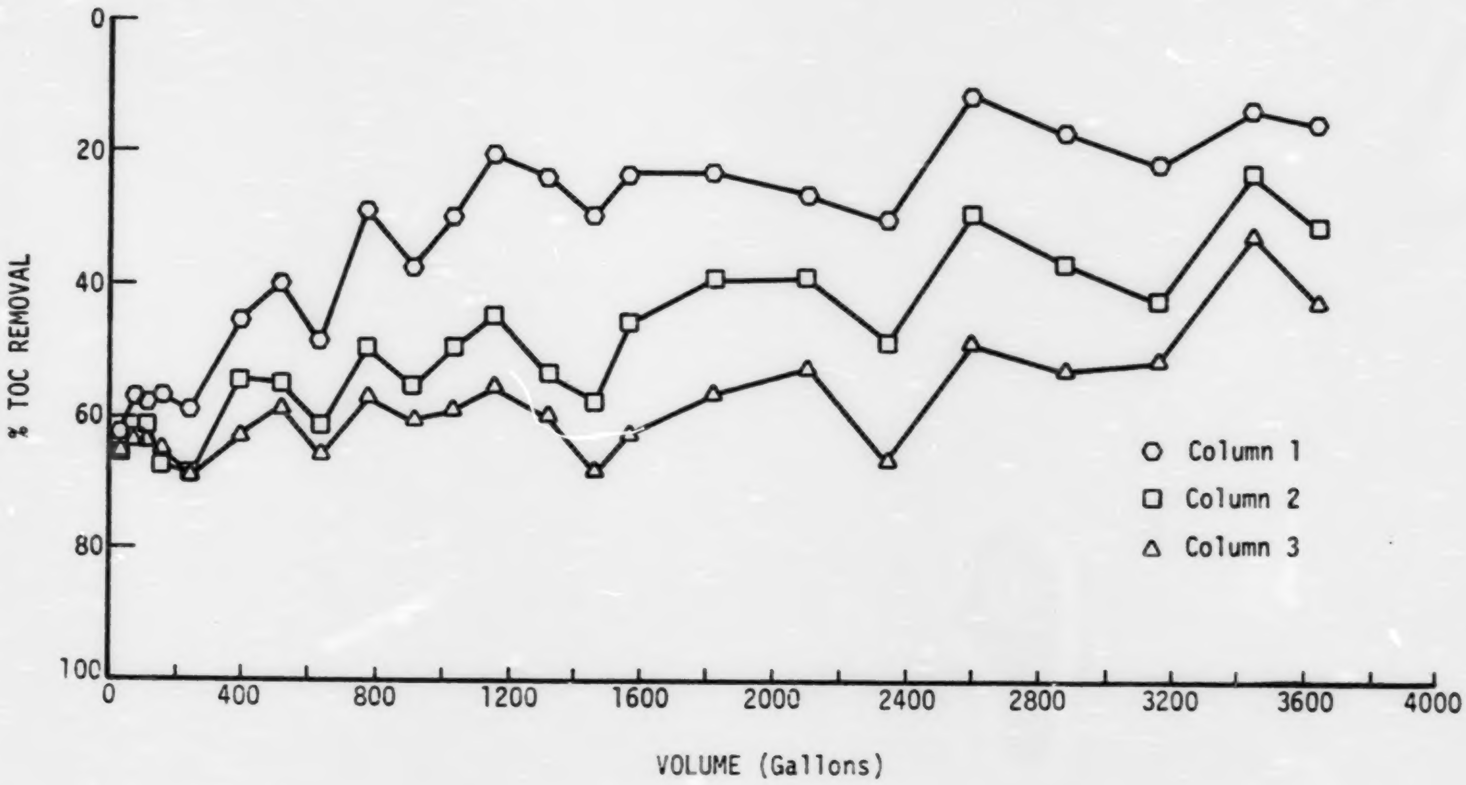


FIGURE V-17

% TOC REMOVAL VS. VOLUME THROUGHPUT - PHYSICAL-CHEMICAL TREATMENT

(COLUMNS 1, 2 & 3)



loading standpoint, adsorption is attractive in that at exhaustion a COD loading of approximately 0.8 lbs COD/lb carbon can be achieved, as seen in Figure V-14. This not withstanding, effluent quality in this case would limit the usage of the physical-chemical treatment for wastewater recirculation because of the presence of non-adsorbable materials.

Tertiary Carbon Treatment

In the tertiary carbon treatment approach as applied to the Anderson Plant, the envisioned process would consist of the existing biological treatment system followed by filtration and granular activated carbon adsorption. This treatment train would remove both the adsorbable, non-biodegradable organics (with carbon) and the less adsorbable, highly biodegradable organics (with activated sludge).

Adsorption Isotherms (1973-1974)

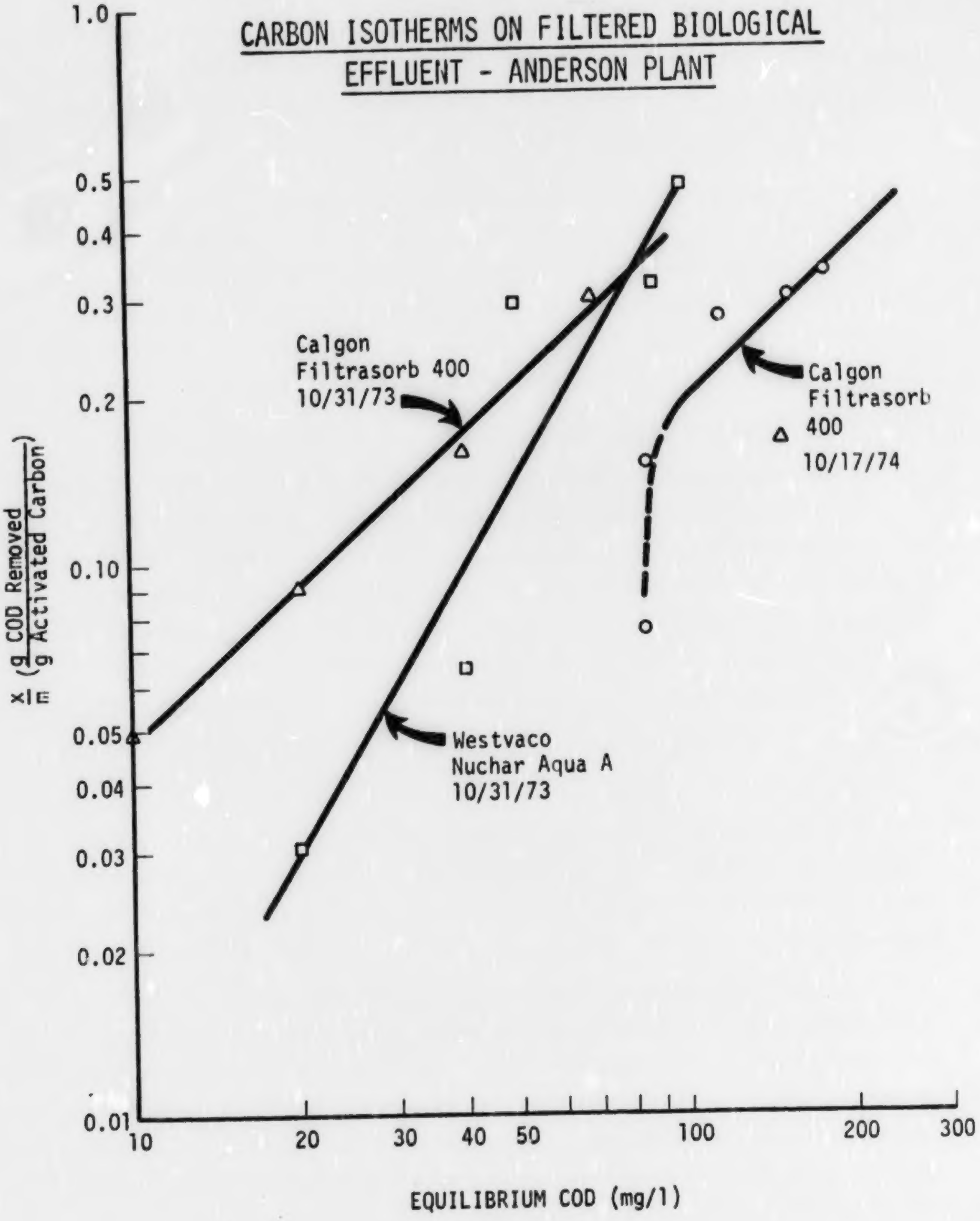
Adsorption isotherms were completed on the filtered and coagulated, filtered biological effluent in a manner similiar to that described previously. Figure V-18 shows the isotherms for the filtered effluent while Figure V-19 shows the isotherms for the coagulated, filtered effluent. In both cases, the carbon isotherms are relatively logarithmic and show acceptable COD removals at moderate carbon loadings. At low carbon loadings, equilibrium COD levels appear to be 30-50 mg/l, which compare favorably with the column study results discussed below.

Carbon Column Studies (1973-1974)

Two distinct carbon column studies were run at the Anderson Plant, the first in December, 1973 and the second in January, 1974. The results of the December test are shown in Figures V-20 through V-26 while the January test results are shown in Figures V-27 through V-29.

During the December column tests, effluent COD levels of 20-50 mg/l were observed at removal efficiencies of approximately 80-85%. Correspondingly, COD loadings as high as 0.30 lb COD/lb carbon were observed, although the column test was terminated before exhaustion because of plugging problems. Column influent was settled biological effluent and thus suspended and colloidal material eventually plugged the columns.

FIGURE V-18



CARBON ISOTHERMS ON FILTERED BIOLOGICAL EFFLUENT - ANDERSON PLANT

FIGURE V-19

CARBON ISOTHERM ON COAGULATED, FILTERED BIOLOGICAL EFFLUENT - ANDERSON PLANT

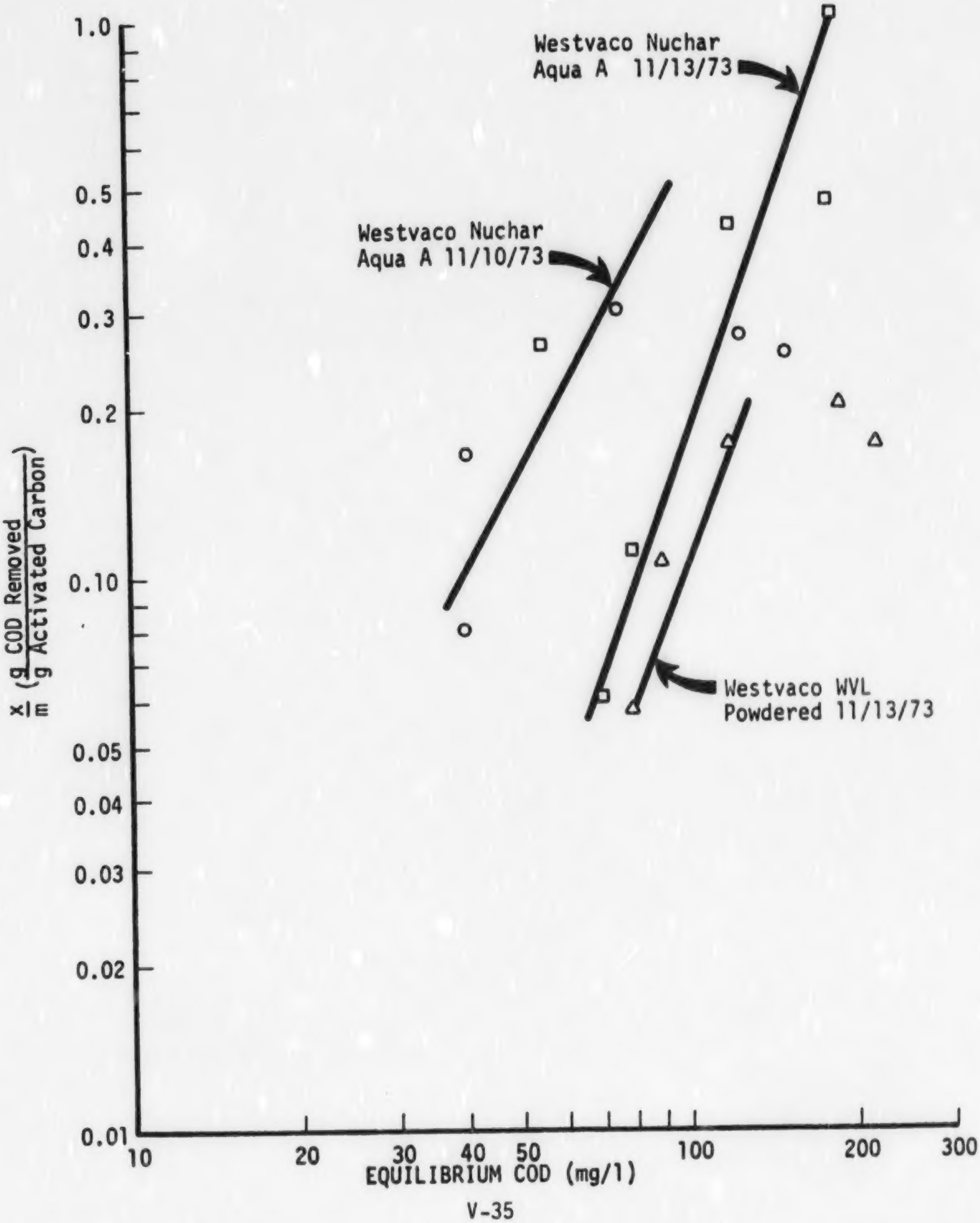


FIGURE V-20

COD VS. VOLUME THROUGHPUT - BIO-EFFLUENT (COLUMNS 1,2 & 3)

DECEMBER 1973

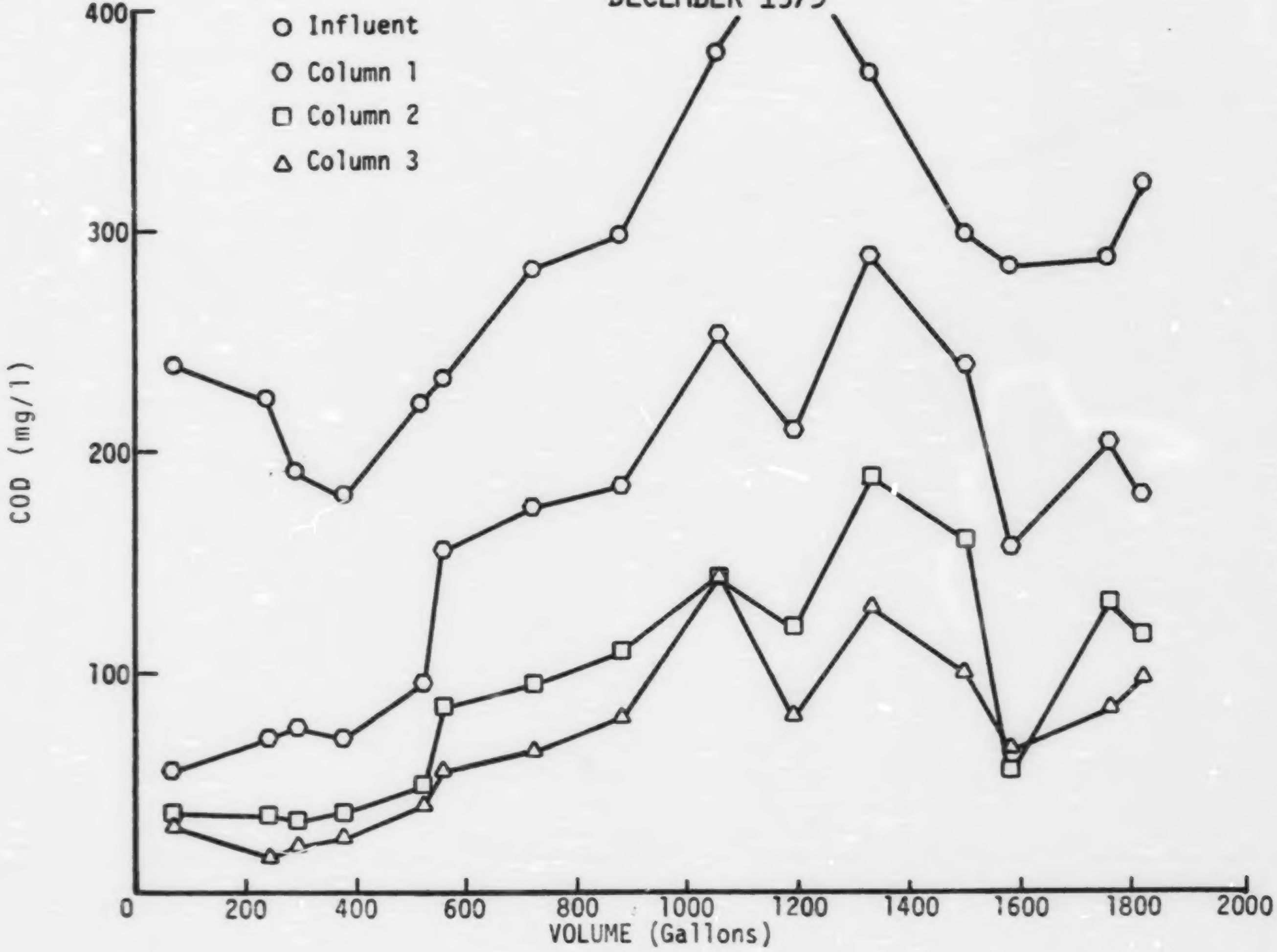


FIGURE V-21

COD VS. VOLUME THROUGHPUT - BIO-EFFLUENT (COLUMNS 4,5 & 6)

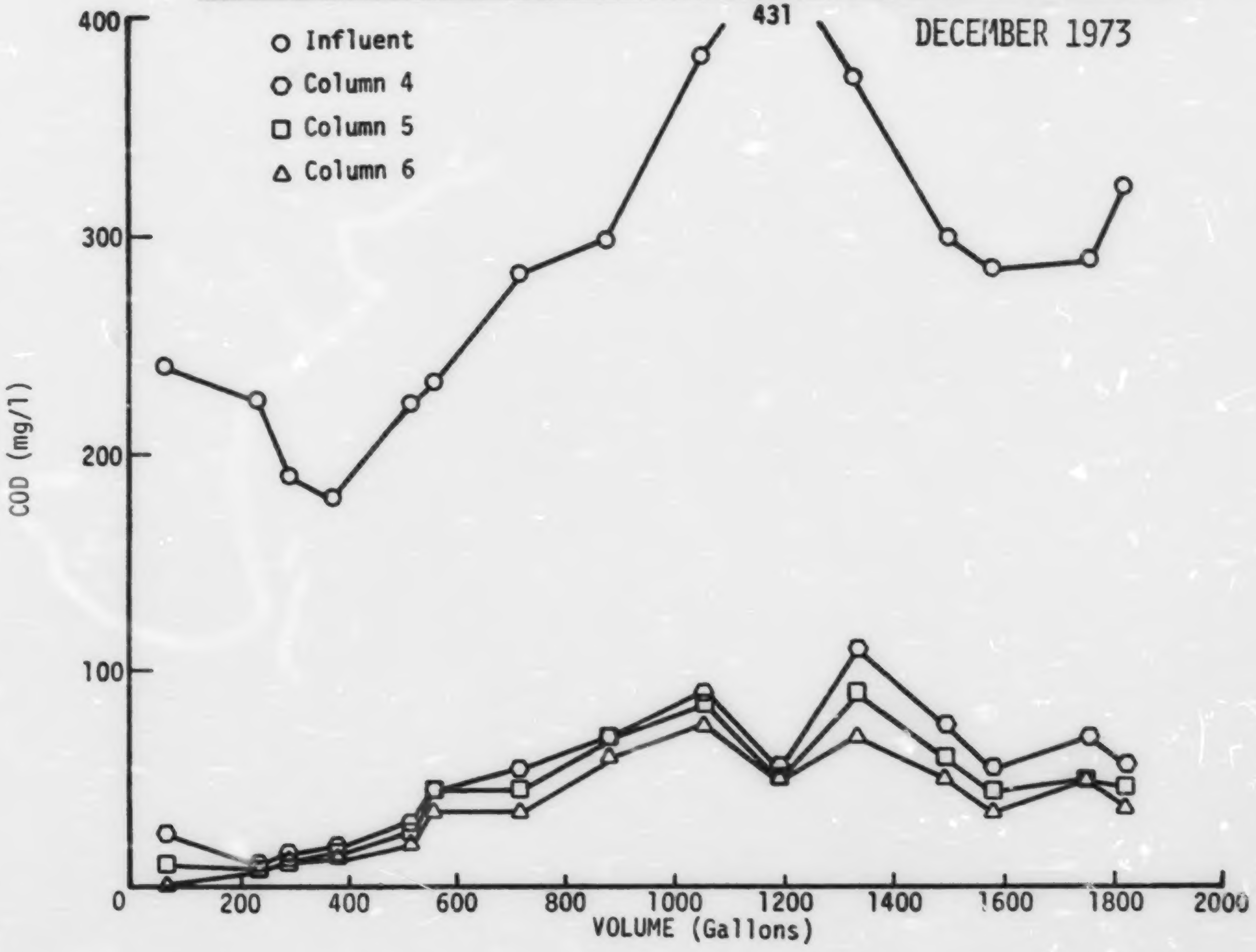
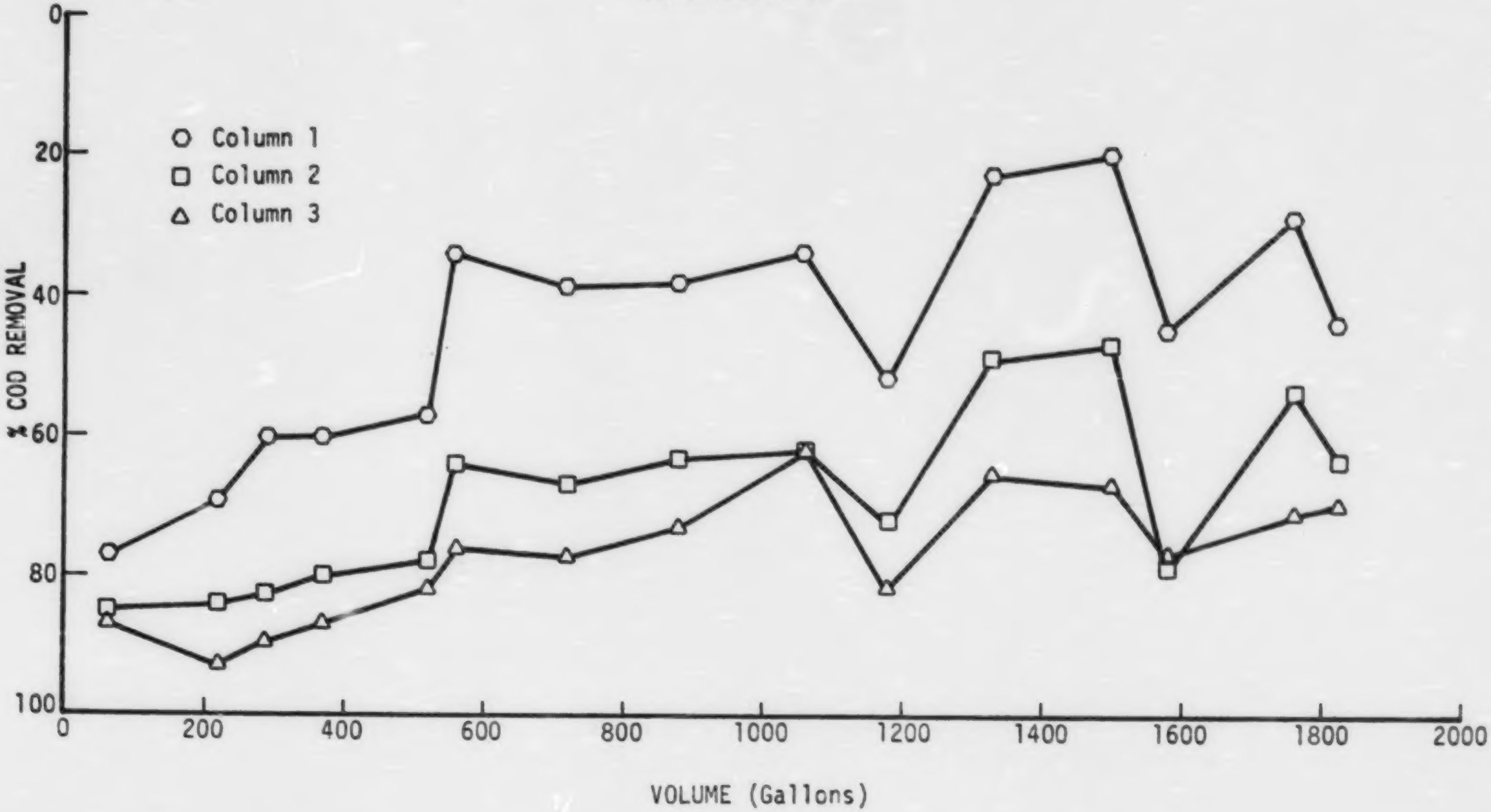

DECEMBER 1973

FIGURE V-22

% COD REMOVAL VS. VOLUME THROUGHPUT - BIO-EFFLUENT

(COLUMNS 1,2 &3)

DECEMBER 1973



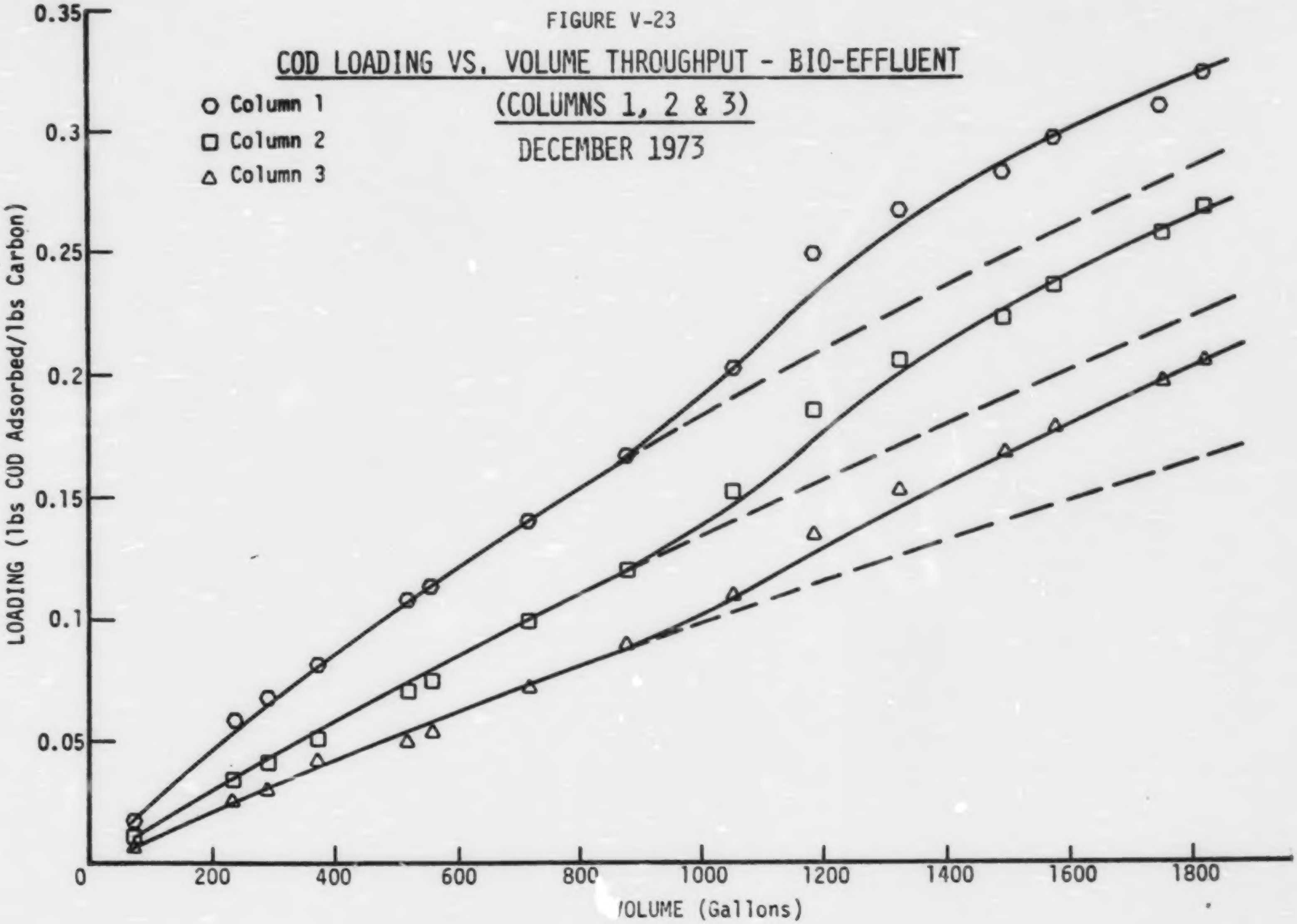


FIGURE V-23

COD LOADING VS. VOLUME THROUGHPUT - BIO-EFFLUENT

(COLUMNS 1, 2 & 3)

DECEMBER 1973

FIGURE V-24

TOC VS. VOLUME THROUGHPUT - BIO-EFFLUENT

(COLUMNS 1, 2 & 3)

DECEMBER 1973

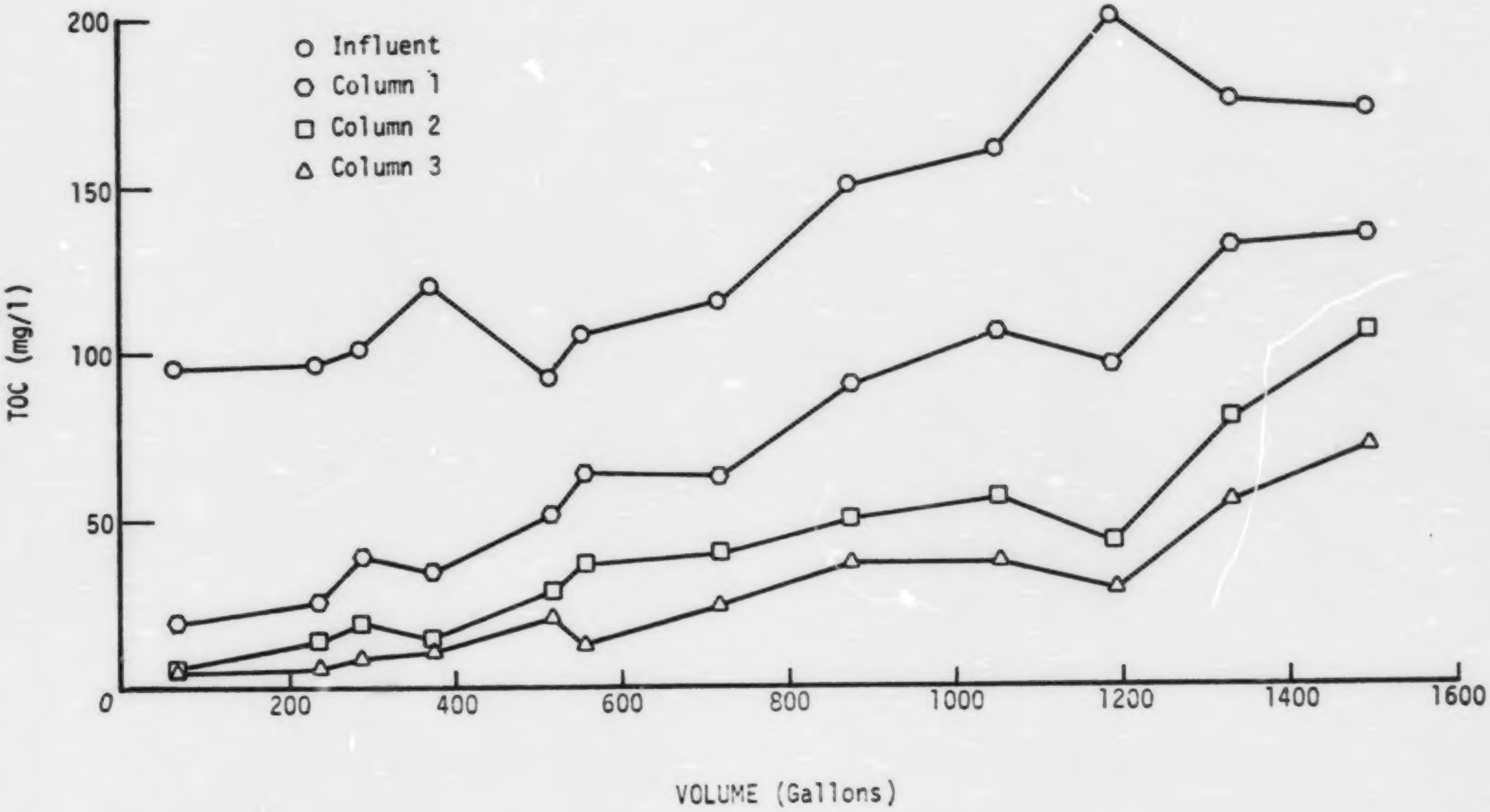


FIGURE V-25

TOC VS. VOLUME THROUGHPUT - BIO-EFFLUENT

(COLUMNS 4, 5 & 6)

DECEMBER 1973

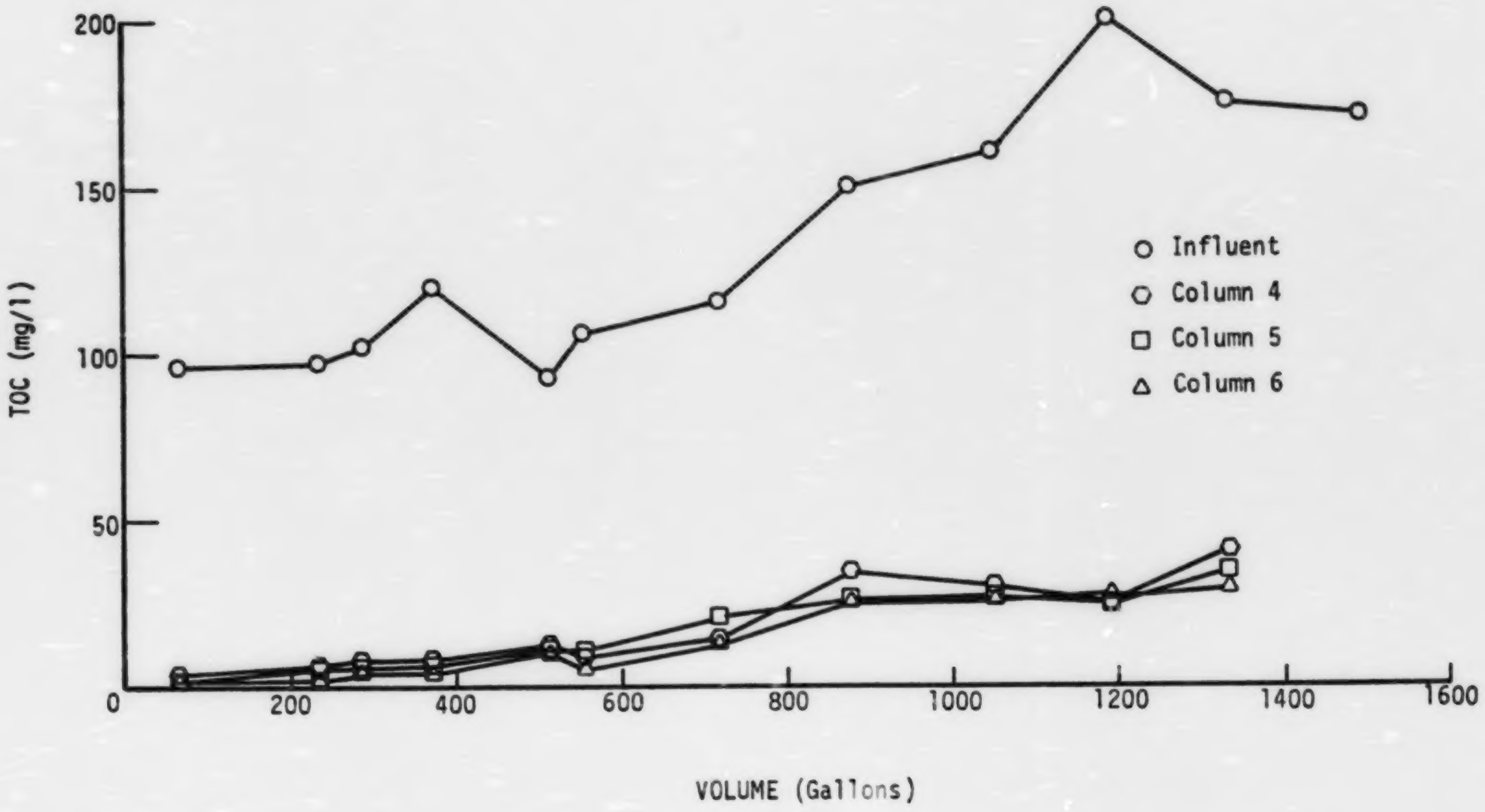


FIGURE V-26

% TOC REMOVAL VS. VOLU[illegible] THROUGHPUT - BIO-EFFLUENT

(COL[illegible] 1, 2 & 3)

DECEMBER 1973

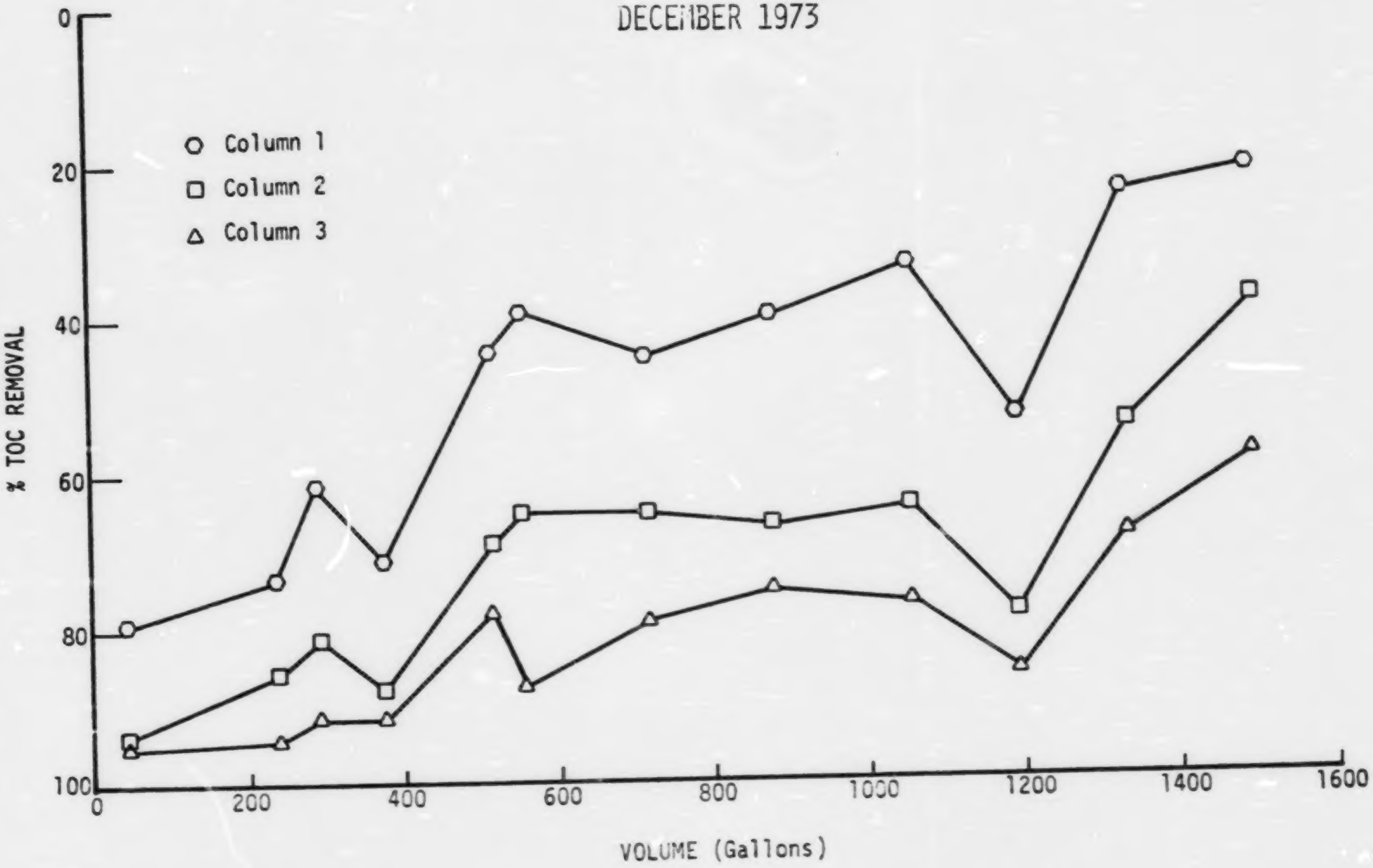


FIGURE V-27

COD VS. VOLUME THROUGHOUT - BIO-EFFLUENT JANUARY 1974

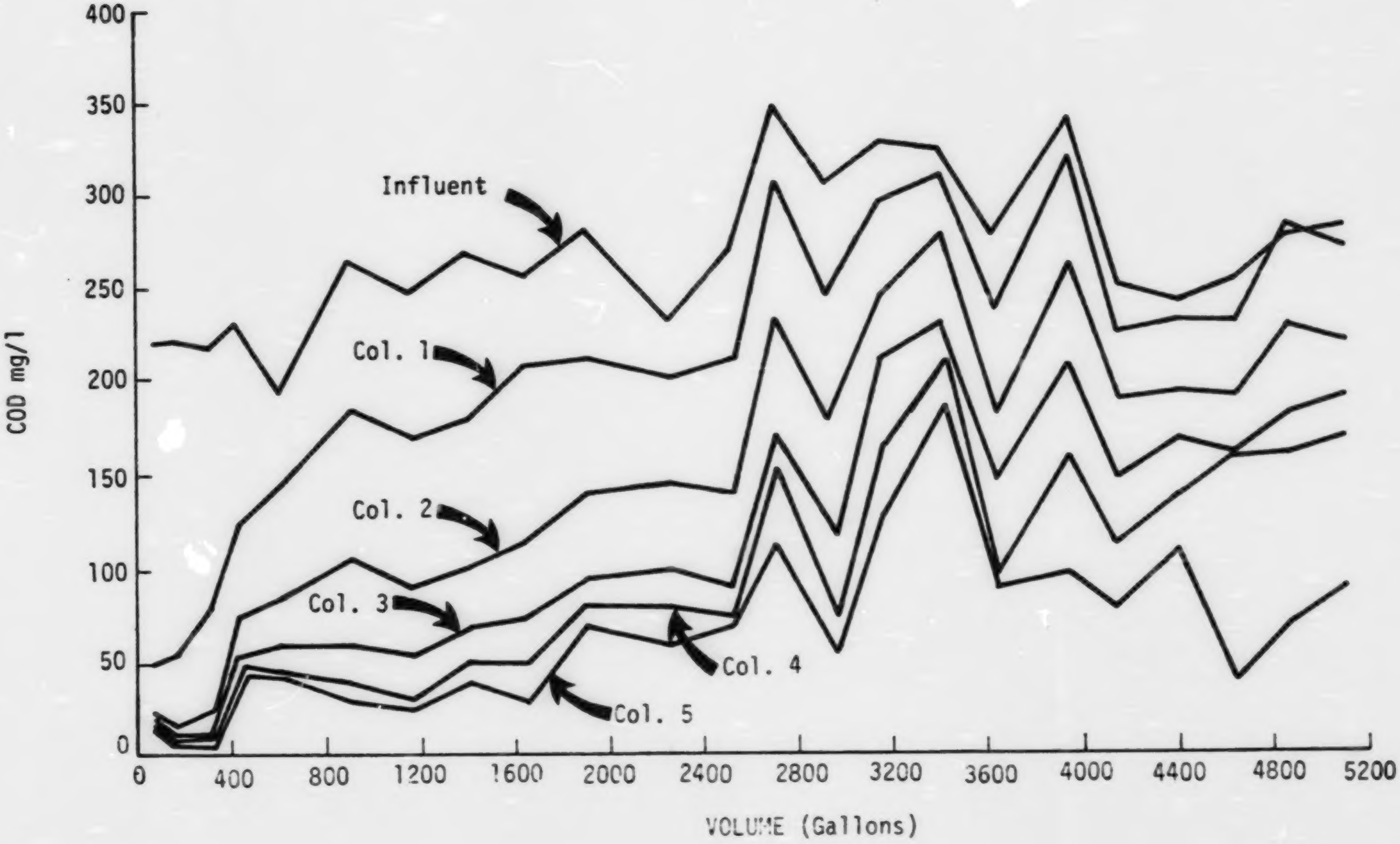
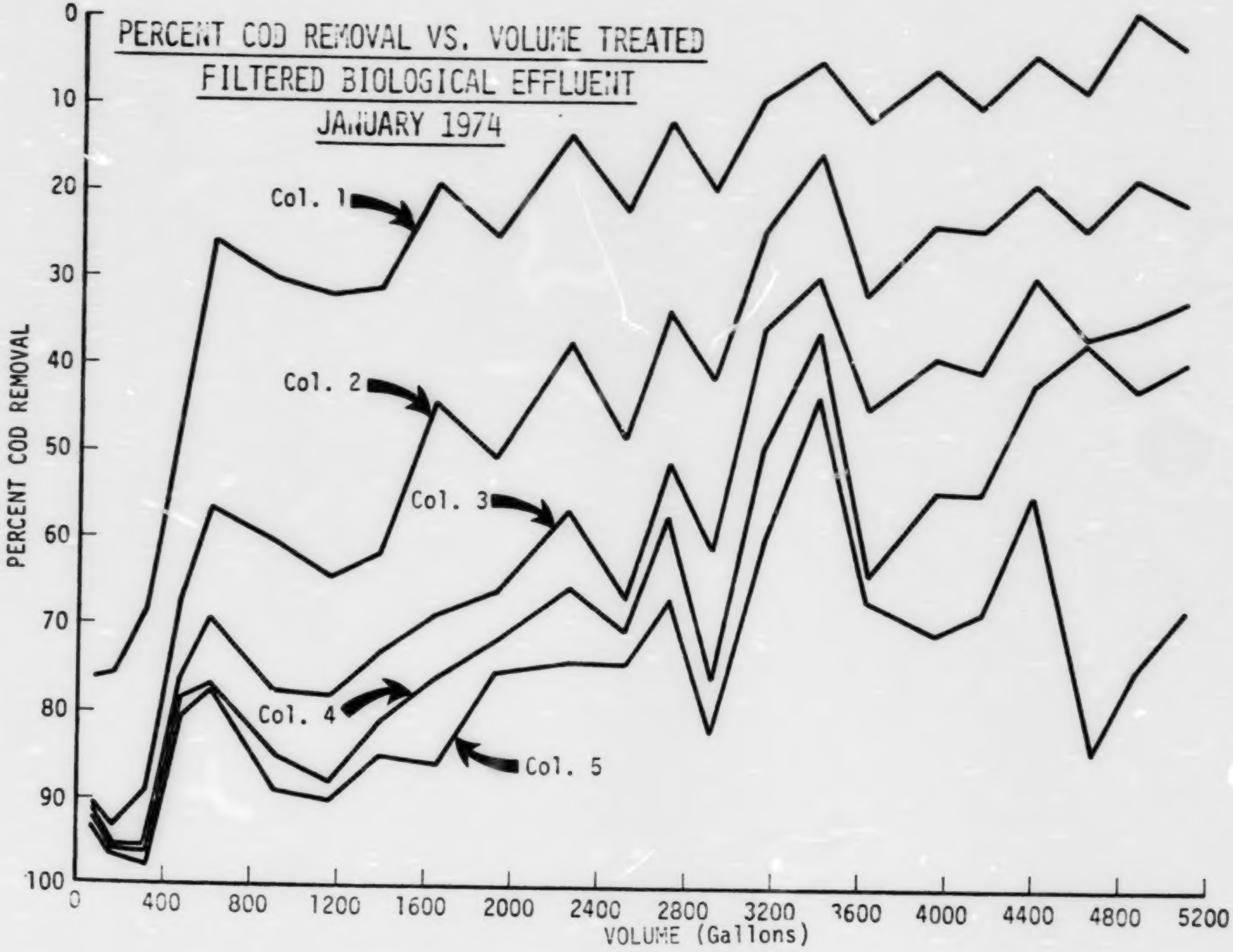


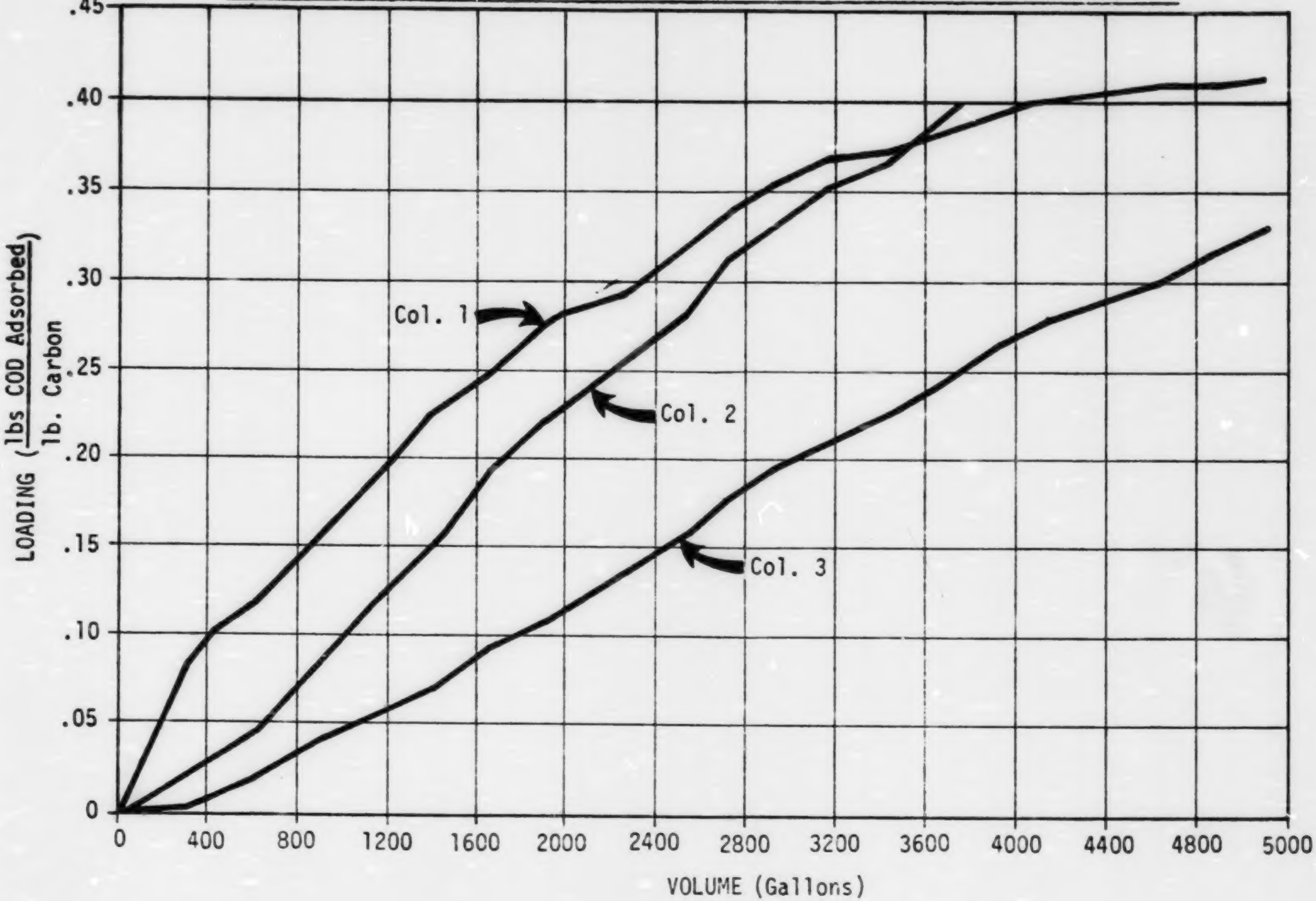
FIGURE V-28



PERCENT COD REMOVAL VS. VOLUME TREATED
FILTERED BIOLOGICAL EFFLUENT
JANUARY 1974

FIGURE V-29

COD LOADING VS. VOLUME THROUGHPUT BIO-EFFLUENT JANUARY 1974



The test completed in January was conducted utilizing filtered biological effluent in order to improve the column performance. The results of this test, shown in Figures V-27 through V-29, are comparable to the December test in terms of COD removal efficiency although significantly higher carbon loadings were observed because the columns were carried to exhaustion. On the basis of this test, the carbon loading at exhaustion approaches 0.40 lbs COD/lb carbon, which is within an acceptable range for design.

Tertiary Pilot Plant (1975-1976)

A tertiary pilot plant was operated at the Anderson Plant by OCF environmental personnel in conjunction with the pilot cooling loop experiments. Pilot units were not operated for treatability purposes, but rather to supply the cooling loops with high quality reclaimed makeup wastewater. The treatment scheme consisted of upflow sand filtration, activated carbon adsorption, and disinfection.

Several operational difficulties were observed during the experiment; the most serious problem was plugging of the carbon column with suspended and colloidal solids. This was most likely a result of the absence of any backwash or air scour capabilities. The plugging phenomenon underscores the need for sand filtration prior to adsorption and for close control of coagulant addition practices. Several disinfection agents were tested: ozone, sodium hypochlorite, and gaseous chlorine. Both ozone and sodium hypochlorite caused oxidation of dissolved iron (residual from $FeCl_3$ addition in the treatment plant) and the resultant precipitation of $Fe(OH)_3$; the sludge settled in the cooling system storage reservoir. While this oxidation - precipitation process was not observed when chlorine was used as the disinfection agent, the same process should have occurred. Successful operation of the reclaimed wastewater recycle scheme will require close control of ferric chloride addition so as to prevent this oxidation - precipitation.

Pilot plant performance is summarized in Table V-5. It is obvious that the three tertiary treatment processes will provide an effluent which is relatively free of suspended solids and organic contaminants. Equivalent or even better performances may be expected with a full scale system.

TABLE V-5

PILOT PLANT PERFORMANCE SUMMARY*

Parameter	Secondary Clarifier Effluent	Sand Filter Effluent	Carbon Column Effluent	Disinfected Effluent
TOC (mg/1)	28	29	18	18
BOD_5 (mg/1)	10	9	7	5
TSS (mg/1)	3	4	2	3
NH_3-N (mg/1)	2.2	-	1.9	1.8
Cr^T (mg/1)	0.10	-	0.05	0.04
Oil & Grease (mg/1)	2.6	2.9	2.0	1.8
Phenol (µg/1)	2.9	-	0.7	1.1

*Mean values - 9/7/75 to 3/10/76.

Adsorption Isotherms (1976)

Adsorption isotherm tests were conducted on filtered secondary clarifier effluent and filtered pilot cooling loop blowdown in April, 1976. These studies were performed for three reasons:

- to determine if the adsorption characteristics of the Anderson wastewater had changed from that exhibited in 1973-1974
- to investigate the adsorbability of contaminants in the pilot cooling loop blowdown
- to correlate carbon loadings based on COD with those based upon TOC

The tests were conducted in a manner similar to that described previously. Results are presented in the adsorption isotherms, Figures V-30 and V-31.

As shown in Figure V-30, the adsorption characteristics of the treatment plant effluent have not changed to any great degree since 1973-1974. Both sets of isotherms indicate that ridiculously low carbon loadings would be required to reduce pilot cooling loop waters to equilibrium levels attained through adsorption of secondary effluent, because the cooling loop waters had already been subjected to adsorption prior to use as makeup in the loop.

According to Figure V-30, at a carbon loading of 0.40 lb COD/lb carbon, secondary effluent COD would be reduced to approximately 90-100 mg/l. In order to interpret this loading in terms of TOC, least-squares linear regression analyses were performed on the concentrations measured during the isotherm tests. Secondary effluent COD was found to be related to TOC as follows:

$$\text{COD} = 4.2\,(\text{TOC}) + 39$$

$$\text{Correlation Coefficient} = r = 0.95$$

Applying the equation to the COD values results in TOC concentrations of 12 - 15 mg/l and carbon loadings of 0.03 - 0.05 lb TOC/lb carbon.

CARBON REGENERATION

One advantage gained through the use of granular carbon for adsorption rather than powdered carbon is that granular carbon can be recovered and regenerated by heating. A percentage of carbon is lost due to

FIGURE V-30

CARBON ISOTHERMS - COD

April, 1976

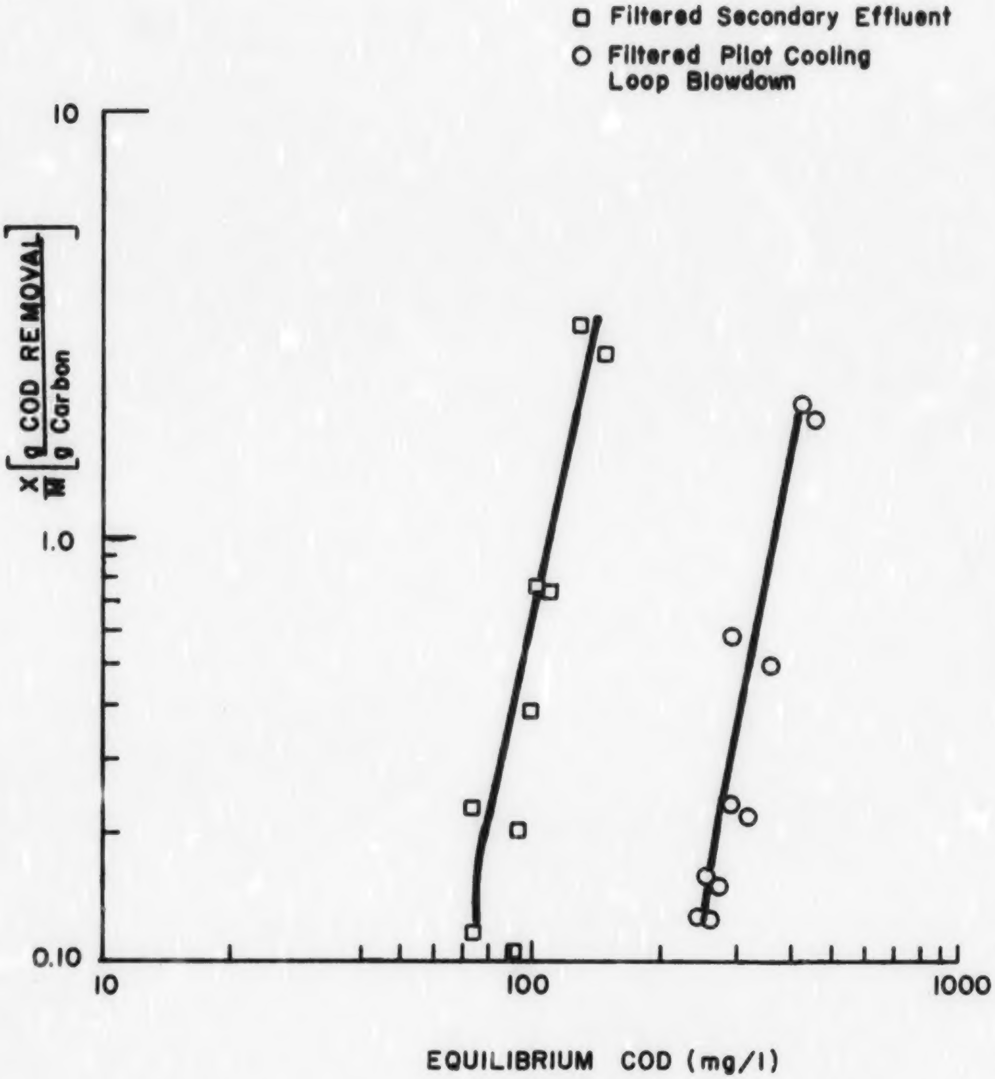
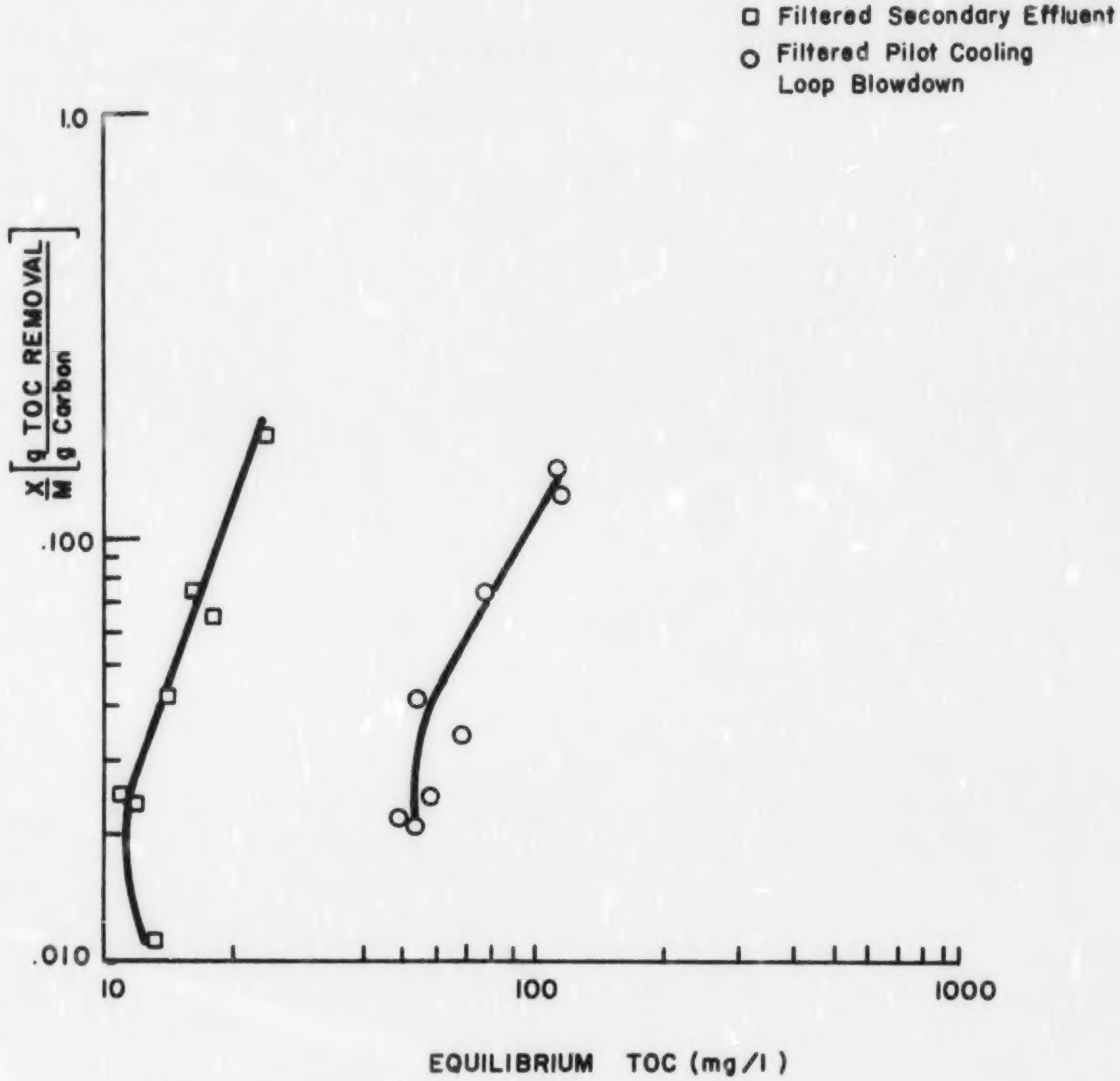


FIGURE V-30

CARBON ISOTHERMS-TOC
April, 1976

FIGURE V-32



physical breakdown during transport and combustion during regeneration. Also, a certain percentage of the adsorption capacity of carbon cannot be regained through regeneration. While physical carbon losses cannot be estimated on the basis of test results, the decrease in capacity following regeneration can be determined through laboratory regeneration studies.

Two samples of carbon, one virgin and one exhausted (from the pilot carbon adsorption column) were delivered to the Westvaco Co. laboratory in Covington, Virginia. The apparent density and capacity (as expressed by the iodine number) were measured for virgin, spent, and regenerated carbon. Regeneration was conducted for 30 minutes at 1000° C. As shown in Table V-6, density recovery was 98% and iodine number recovery was 98.4%, resulting in a net carbon capacity regeneration recovery of 96.4%.

TABLE V-6

CARBON REGENERATION RESULTS

Carbon	Apparent Density (gm/cm^3)	Iodine Number
Virgin	0.490	1033
Spent	0.565	686
Regenerated	0.480	1017

OZONATION

Bench scale ozonation tests were run on samples of effluent from the pilot carbon columns (which provided makeup to the pilot recirculation process cooling loop) to determine the effectiveness of this method of disinfection. Disinfection will be necessary for reuse of any reclaimed wastewater to limit microbial growth in the cooling systems and for the obvious public health reasons. Ozonation will not increase the dissolved solids concentration of the wastewater.

Procedure

A laboratory scale ozone generator operating on pure oxygen was connected through a rotameter to diffusers in two glass contact chambers (each 3 ℓ capacity) in series. The system was calibrated before and after the sample runs by filling both contact chambers with KI solutions and running the generator for three minutes at a specific power output and gas flow. Duplicate sample runs were made at identical settings with 3 liters of KI solution in the second contact chamber. The KI solutions were then titrated with PAO to determine the amount of O_3 absorbed.

Data Analysis

One of the samples was analyzed before and after ozone contact for COD, TOC, and fecal coliform concentrations. The results shown in Table V-7 indicate that a relatively low dosage of ozone had little effect on COD and TOC but was effective in eliminating fecal coliforms. While this test is not rigorous enough to provide design data, it demonstrates that there is no serious interference (by COD or TOC) with the disinfecting action of ozone.

TABLE V-7

OZONATION RESULTS

O_3 dose measured in calibration solution of KI	9.45 mg/l
O_3 dose measured in KI solution downstream of sample	8.85 mg/l
O_3 dose for sample (calculated)	0.60 mg/l
COD before contact	61 mg/l
COD after contact	55 mg/l
COD reduction	6 mg/l
Percent COD reduction	10%
TOC before contact	41 mg/l
TOC after contact	41 mg/l
TOC reduction	0%
Fecal Coliforms* before contact	63/100 ml
Fecal Coliforms* after contact	0/100 ml
Fecal Coliform Kill	100%

*Average value of three determinations on sample with membrane filter technique.

CHLORINATION

Chlorination is another alternative process for disinfection. Bench scale chlorination studies were performed on effluents from both the sand filter and the carbon columns used in the treatability studies. Test procedures consisted of adding a measured dosage of liquid bleach (known chlorine content) to a 500 ml sample, mixing the solution gently for 30 minutes, and analyzing the liquid for residual chlorine and fecal coliform concentrations. Chlorine dosages and chlorine residuals were found to be related as depicted in Figure V-32. Sand filter effluent exhibited a significant chlorine demand. Results listed in Table V-8 show that a chlorine dosage of 1 - 2 mg/l Cl_2 will result in a residual of 1 mg/l Cl_2 and virtually eliminate all fecal coliform organisms from an activated carbon adsorption effluent.

ION EXCHANGE

Ion Exchange may be required for removal of total dissolved solids or constituents such as hardness, sulfate, silica, and zinc from the reclaimed wastewater. Ion exchange units could be operated on a side stream around recirculating cooling systems or at the treatment facility. Laboratory scale ion exchange columns were operated at the Anderson Plant to determine which, if any, of the pilot carbon column effluent inorganic TDS constituents would be difficult to remove.

Procedure

A set of four 1 inch I.D. 5 foot long plexiglass columns were charged with samples of three Rohm & Haas resins, and fed with positive displacement variable speed pumps from the steel tank containing carbon column effluent, in use as a reservoir for makeup water to the pilot process cooling system. The piping system was 1/4 inch plastic or rubber tubing and was arranged so as to allow series or parallel operation of backwash capability. Column 1 was charged with 308 ml of MB-1 resin (a mixed-bed combination of IRA-120 and IRA-400), Column 2 was left empty to receive the anionic component of the MB-1 (IRA-400) during the regeneration process, and Columns 3 and 4 were charged with 308 ml of IRA-93 (strong anionic) and 334 ml of Amberlite 200 (strong cationic) resins respectively, and run in series. All resins used were in either the hydrogen or the hydroxide form.

TABLE V-8

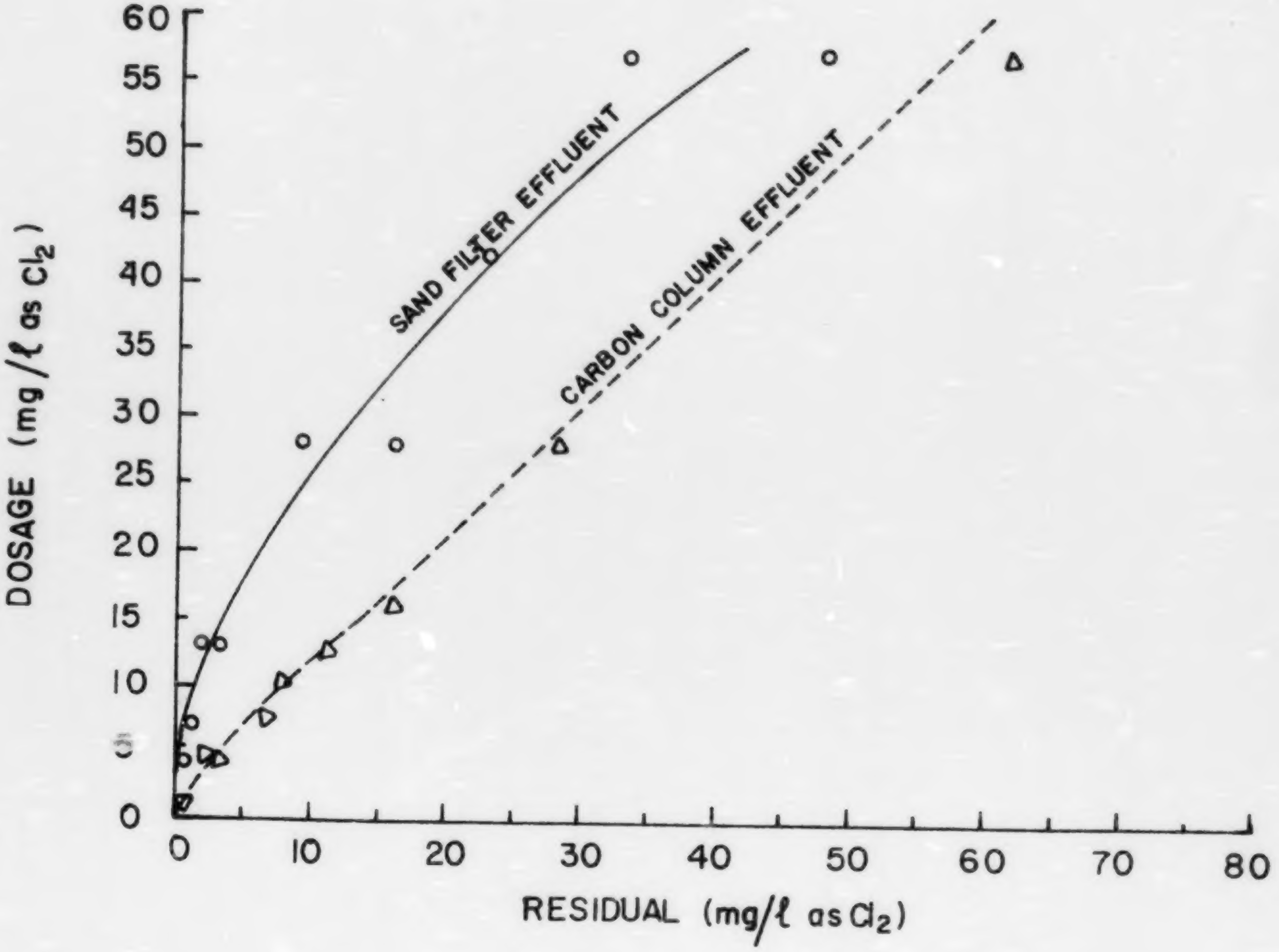
CARBON EFFLUENT CHLORINATION RESULTS

Sample	Chlorine Dosage $(mg/l\ Cl_2)$	Chlorine Residual after 30 min. $(mg/l\ Cl_2)$	Coliform Concentration (organisms/100 ml)
6	0	0	TNTC
7	1.4	0.4	0
8	4.5	2.4	108
9	13	11.3	144
10	28	28.2	4
11	57	62	0
12	1.4	0.2	2
13	4.5	3.3	0
14	7.5	6.9	0
15	10.5	7.9	0
16	13	11.0	4
17	16.3	16.0	0

*Tests performed 1/23/74.

FIGURE V-32

CHLORINE DEMANDS



The columns were exhausted by feeding at an average rate of 42 ml/min, or 0.13 bed volumes per minute (0.13 B.V./min). Before a run, each column was backwashed at a rate sufficient to provide 50% expansion of the bed, and then allowed to settle. After the MB-1 resin was exhausted, it was backwashed at a rate high enough to wash all of the IRA-400 resin out and into the receiving Column 2, where it was regenerated. The sample of Amberlite 200 was received in the sodium form and was regenerated before it was tested. Regeneration of anionic resins was accomplished by feeding 2 liters of 4% NaOH solution per 300 ml resin at a rate of 30 ml/min, or 0.1 BV/min. Regeneration of cationic resins was done by feeding 4 liters of 5% H_2SO_4 solution per 300 ml of resin at 30 ml/min. During each resin exhaustion run the column effluent was sampled hourly and analyzed for conductivity, silica, pH, sulfate, total carbon, total dissolved solids, alkalinity, calcium hardness, and total hardness.

Data Analysis

The MB-1 resin was exhausted once, regenerated, and then exhausted again. The effluent analysis data plotted against test duration resulted in the breakthrough curves shown in Figure V-33 and V-34. This resin removed all ionic constituents of the wastewater down to concentrations that were beyond the limits of detection for the analysis procedures used. Only 80% removal of total carbon was achieved at the beginning of the runs (to 10 mg/l), and the effluent carbon content began to rise immediately. This is to be expected as non-ionic organics would be removed only by secondary adsorption or filtration mechanisms in the column. Total dissolved solids decreased from 340 to 80 mg/l on passage through the column, indicating that approximately 80 mg/l of organics and non-ionic species are present in the carbon column effluent.

Breakthrough, as indicated by conductivity and TDS, occurred at 19 hours for the virgin MG-1 resin and at 10 hours for the regenerated resin, indicating that regeneration was not complete. The pH data (not presented here) showed that while the effluent normally possessed a neutral pH, the effluent from the regenerated resin was very low, indicating that some anionic resin had been lost or was not regenerated during the separation, regeneration, and remixing steps, resulting in

FIGURE V-33

ION EXCHANGE BREAKTHROUGH CURVES
VIRGIN MB - 1 RESIN 10/22/74

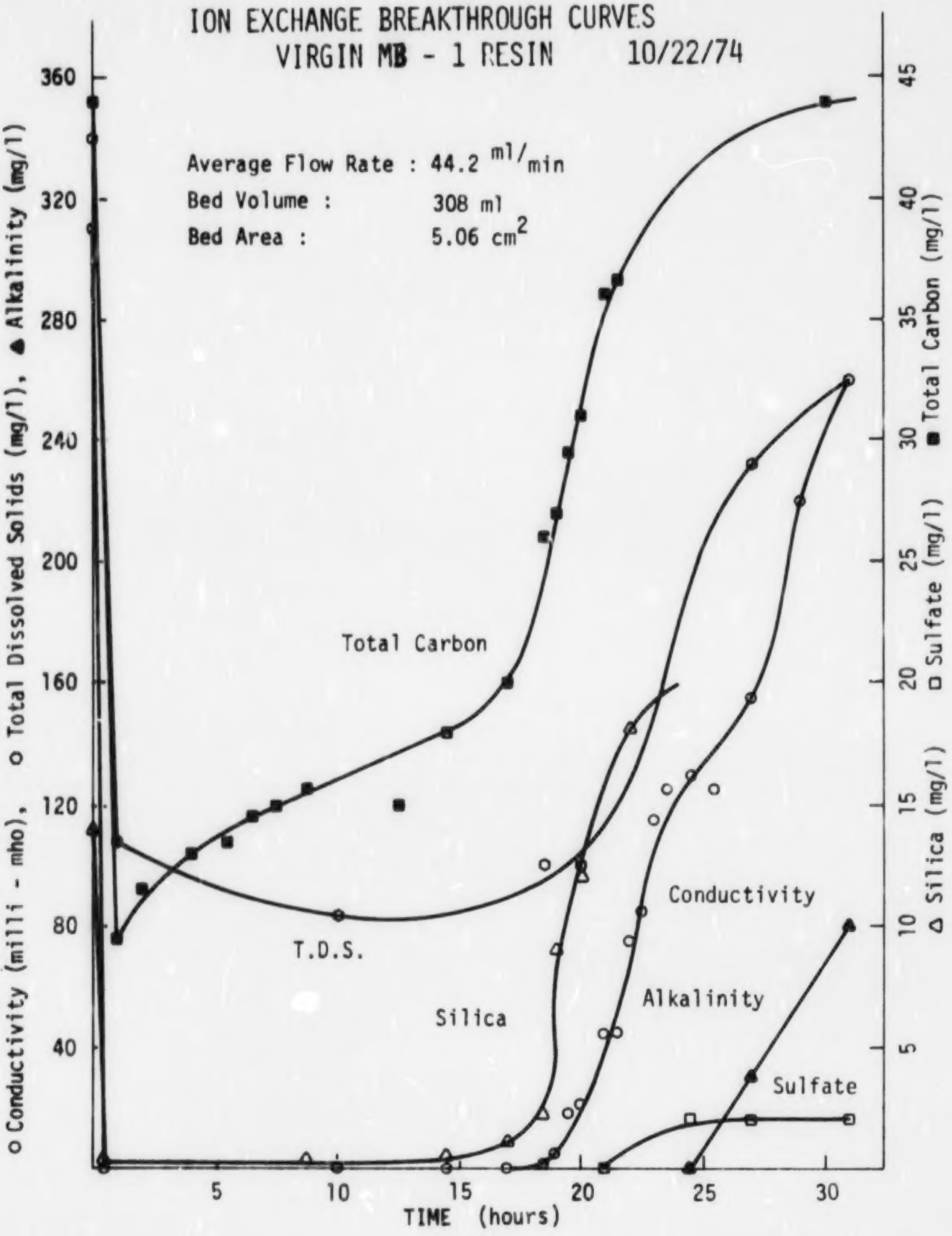
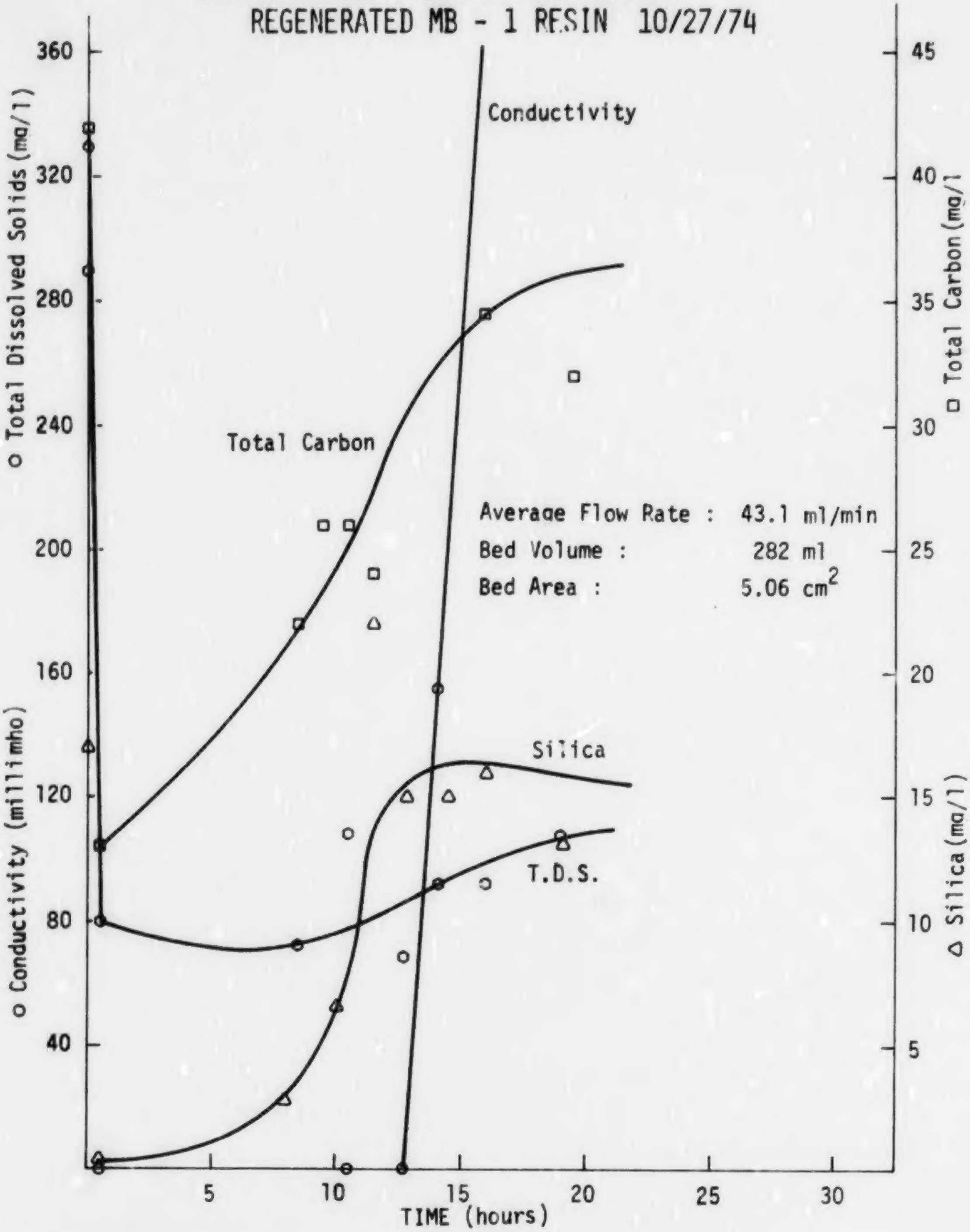


FIGURE V-34

ION EXCHANGE BREAKTHROUGH CURVES
REGENERATED MB - 1 RESIN 10/27/74



a predominantly cationic bed. The performance of the virgin resin is probably closer to that to be expected from a full scale installation. Silica was the first species to exhibit breakthrough, with sulfate following several hours later. Calcium and total hardness never appeared in any of the resin effluents.

Considering the various ionic species present it is estimated that the average equivalent weight of the ionic constituents is 55 mg/meq, so that there are (340-80)/55 meq/liter, or 4.7 meq/liter. The average resin capacity of the MB-1 can be calculated as follows:

$$\text{Capacity} = \left[\frac{19\text{ hr.} + 10\text{ hr.}}{2}\right]\frac{(60\text{ min/hr})(43.6\text{ ml/min})(4.7\text{ meq/l})(1/1000\text{ ml})}{308\text{ ml}}$$

$$= 0.58\text{ meq/ml}$$

This is a rather low value, as a typical resin capacity would be 1.00 to 1.25 meq/ml. The low capacity is due to the necessity of removing silica, which is not easily removed by ion exchange. If effluent hardness is considered as the criterion for breakthrough, the capacity would be above 1.15 meq/l.

The Amberlite 200 and IRA-93 resins were employed in a dual bed arrangement with the cationic Amberlite 200 preceeding the IRA-93. Time constraints prevented the tests from being conducted to breakthrough, but the data showed the removals listed in Table V-9.

TABLE V-9

DUAL BED ION EXCHANGE REMOVALS

Constituent	% Removal	Effluent Level
Conductivity	98%	6 millimho
Silica	50%	7 mg/l
Sulfate	100%	0 mg/l
Total Carbon	50%	20 mg/l
Total Dissolved Solids	86%	60 mg/l
Alkalinity	100%	0 mg/l
Total Hardness	95%	2 mg/l

The capacity of the dual bed resins was at least 0.75 meq/ml with respect to the removals shown above.

Use of either a mixed bed or dual bed ion exchange process will produce an effluent of higher quality than that required with respect to ionic species, but will not remove organic dissolved solids down to the level of TDS in the city water. The capacity of the ion exchange resins tested with respect to silica removal is rather low, requiring frequent regeneration and excess regenerant. The capacity of ion exchange with respect to dissolved solids removal is moderate, being at least 1.2 meq/ml of resin as defined by hardness breakthrough for MB-1 resin.

CHAPTER VI

PILOT PROCESS COOLING LOOP OPERATION AND PERFORMANCE

The process cooling water systems constitute the major use of reclaimed industrial wastewater in this textile fibrous glass total recycle system. Of the estimated 205 gpm of reclaimed wastewater to be generated at the Anderson plant, approximately 44 to 85 percent (during winter and summer loading conditions, respectively), could be utilized in the process cooling water systems. For reclaimed wastewater to be suitable for this use, the dissolved and suspended contaminants present in the wastewater must have no adverse effects on the production processes.

Cooling water recycle trials, the final experiments of Phase I, were designed for determination of any unfavorable effects of reuse. These trials were conducted using an isolated manufacturing process pilot cooling system which operated using reclaimed industrial wastewater. The initial system was designed and constructed between February and August, 1974 and operated from August through November, 1974. Following those trials, due to a significant decrease in manufacturing production level, the cooling water trial was postponed. Additional trials were begun during August, 1975 and continued through May, 1976. This chapter summarizes the results of the pilot cooling trials.

SYSTEM DESCRIPTION

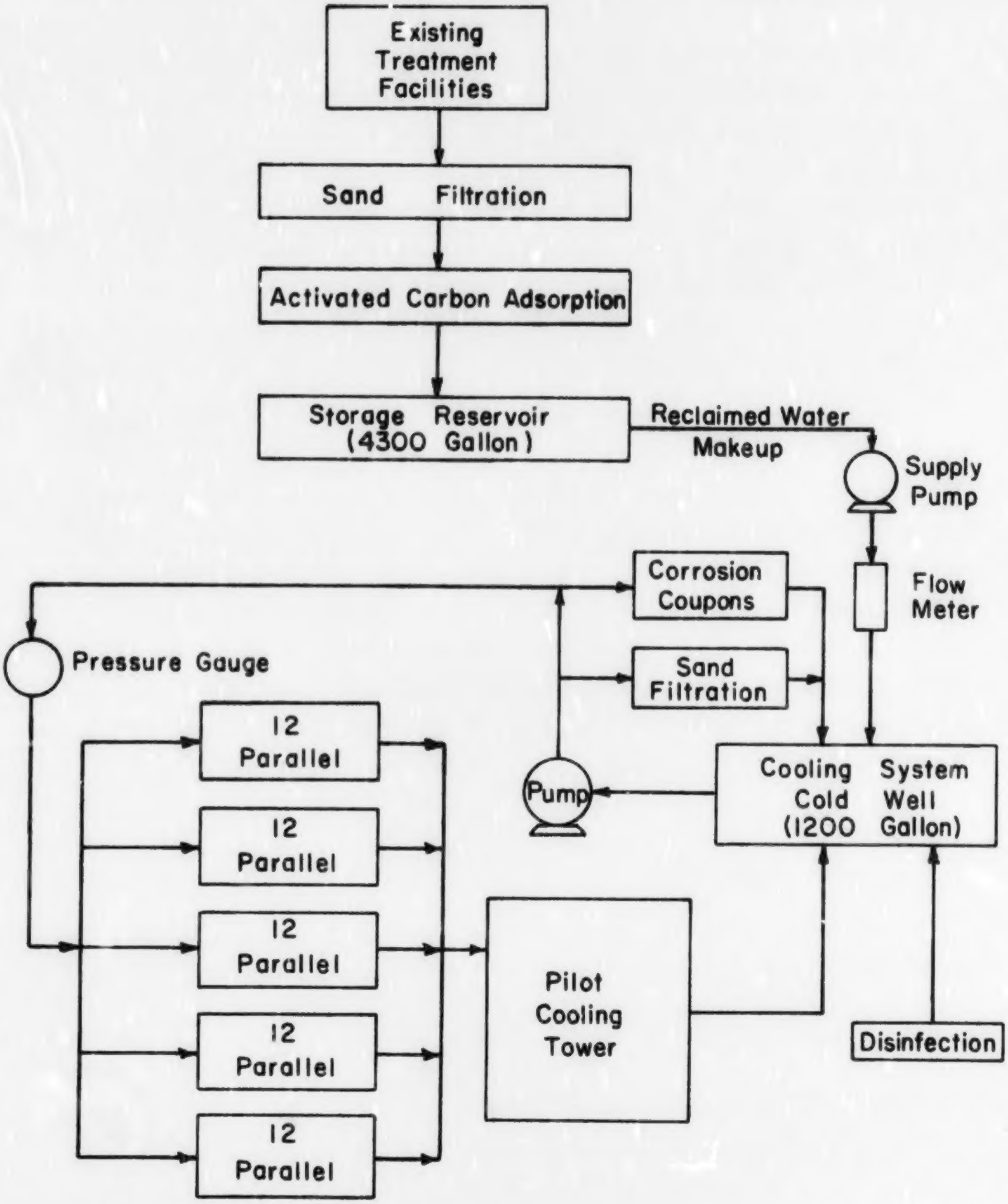
A schematic diagram of the initial pilot reclaimed water supply system and the pilot cooling water system appears in Figure VI-1. The final system as operated during the last 90 days of operation is shown as Figure VI-2.

Wastewater Tertiary Treatment System

Activated sludge secondary effluent was polished by sand filtration followed by activated carbon adsorption and disinfection. The sand filter initially used for tertiary treatment was a gravity flow, rapid sand filter; however, this sand filter was later replaced by an upflow

FIGURE VI-1

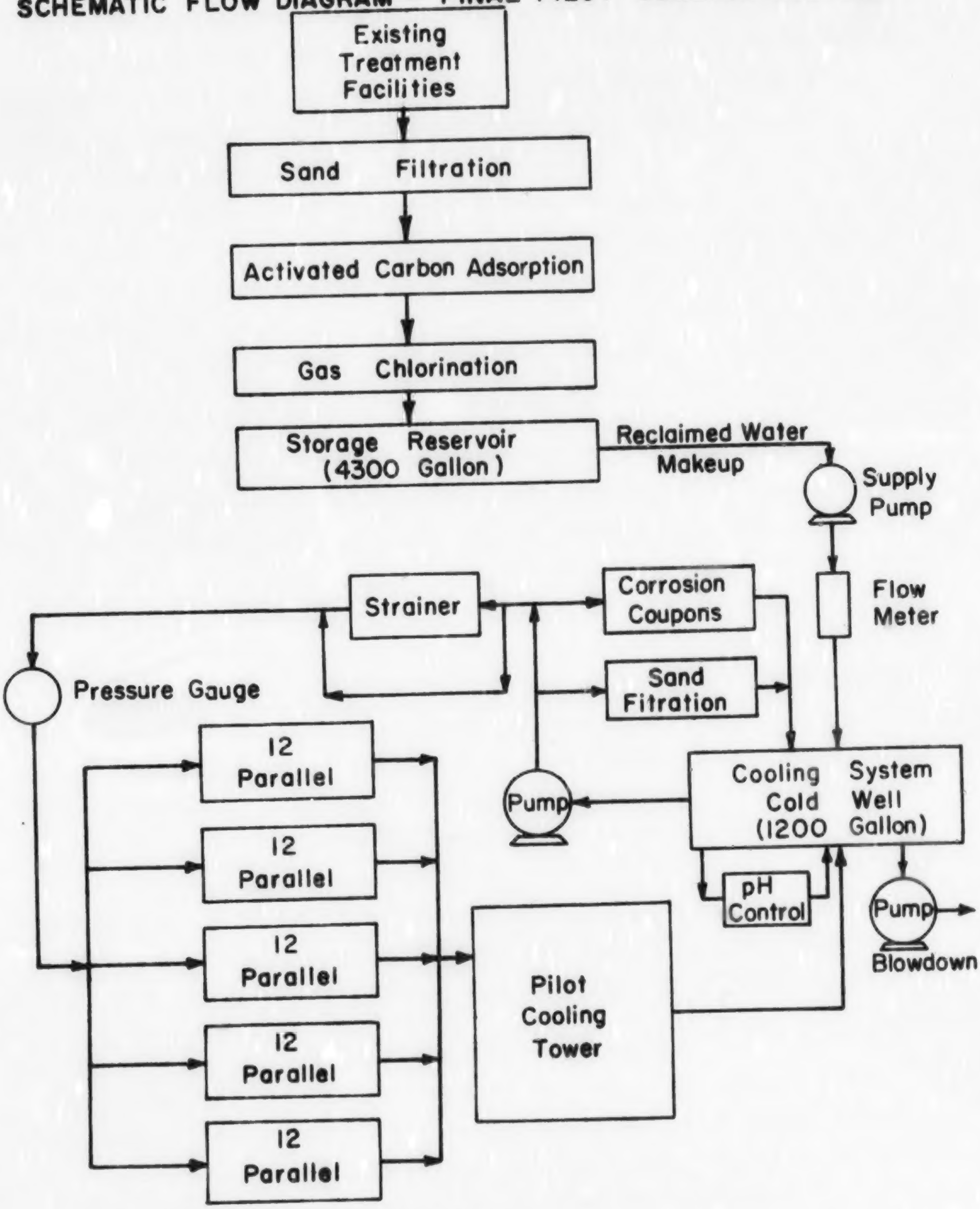
SCHEMATIC FLOW DIAGRAM - INITIAL PILOT COOLING SYSTEM



Five Manufacturing Machines
(Approximately 12 cooling loops per machine)

FIGURE VI-2

SCHEMATIC FLOW DIAGRAM - FINAL PILOT COOLING SYSTEM



Five Manufacturing Machines
(Approximately 12 cooling loops per machine)

pressure sand filter (manufactured by Sea Blue for use in swimming pool water treatment). The activated carbon columns used in both the initial and continuation trials were upflow columns with backwash capabilities. Several means of disinfection were employed during the trials. The types used were: 1) methylene bis-thiocyanate with sodium hypochlorite (added in the cooling loop), 2) combination of methylene bis-thiocyanate, sodium hypochlorite, and a compound containing a quaternary amine and chlorinated phenol (added in the cooling loop), 3) continuous ozonation, 4) sodium hypochlorite only, and, 5) continuous gas chlorination. Gas chlorination of the reclaimed water proved to be the most effective method and was used during the final 90 days of operation.

Cooling Loop

The sidestream sand filter within the pilot cooling loop system was a Sea Blue upflow pressure filter. This filter was operated at several flow rates as shown in Table VI-1 (pg. VI-5). A y-basket strainer was placed in the final cooling system between the supply pump and the cooling loops. The purpose of the strainer was to remove debris from the cooling supply water which could cause plugging in the individual cooling loops. Corrosion coupons of copper, mild steel, aluminum, 304 stainless steel, and galvanized steel were evaluated in the cooling system. The cooling tower used in the cooling system was a Marley Permatower (40 ton). This unit has the capacity to cool approximately 125 gpm of water with an approximate temperature differential of 10°F. An automatic pH controller (manufactured by Uniloc) was installed in the cooling system in fall 1975 to provide continuous pH control in the system. Continuous blowdown from the cooling system was also provided in fall 1975 through installation of a positive displacement pump.

Manufacturing Process Heat Exchangers

The pilot cooling system provided all required cooling for five full-scale glass fiber manufacturing machines. The cooling waters flowed in parallel to the machines; each machine had approximately twelve (12) isolated components that were cooled by twelve separate parallel cooling streams. The pilot trials were operated on manufacturing

TABLE VI-1

DESCRIPTION OF EXPERIMENTS

Experiment Number	Date Begin	Date End	Duration in Days	Operational* Description
1	8-2-74	9-8-74	20	Sidestream Filter 40-45 TPD** Chromate Treatment A
2	9-9-74	9-25-74	16	Sidestream Filter 1 TPD Chromate Treatment A
3	10-10-74	10-24-74	15	Sidestream Filter 1 TPD Polyol Treatment B
4	10-25-74	10-30-74	6	Sidestream Filter 20 TPD Polyol Treatment B
5	10-31-74	11-10-74	11	Sidestream Filter 40 TPD Chromate Treatment C
6	11-11-74	11-22-74	12	Sidestream Filter 1 TPD Chromate Treatment C
7	8-18-75	10-9-75	42	Sidestream Filter 5 TPD Chromate Treatment D
8	11-12-75	1-7-76	56	Continuous blowdown at 10% of makeup rate Sidestream Filter 10 TPD Chromate Treatment E
9	1-27-76	4-25-76	90	Continuous blowdown at 10% of makeup rate Sidestream Filter 10 TPD Chromate Treatment F

*See Table VI-2 for chemical addition descriptions.

**TPD = turnover per day; for example, 2 TPD would indicate that all of the water in the pilot cooling system was filtered twice a day.

positions containing heat exchangers which are subjected to some of the highest heat loadings in the fibrous glass textile operations. Successful operation of the pilot cooling system would therefore indicate the ability to operate the remaining cooling systems using reclaimed industrial wastewater.

Pilot System Monitoring

During the initial pilot trials, the total internal system flow rate, temperature, and pressure were monitored daily. Additionally, one manufacturing position outside of the pilot system was monitored daily for flow rates and temperature differentials.

The reclaimed water makeup and the recirculated cooling water were both analyzed daily throughout the entire experimental period for the following parameters: total organic carbon, suspended solids, dissolved solids (conductivity), total hardness, calcium hardness, silica, alkalinity, chromate, zinc, sulfate, pH, and turbidity. The city water (used as makeup to the other plant cooling systems) and the recirculated cooling water in the plant cooling systems were also analyzed for the same parameters.

SYSTEM PERFORMANCE

The pilot cooling trials conducted under the grant can be divided into two major periods; the initial trials, from August to November, 1974, and the continuation trials, from August, 1975 to May, 1976. Six (6) experiments comprise the ninety (90) days of system operation during the initial trials, and three (3) experiments comprise the 188 days of system operation during the continuation trials. Experimental periods were usually terminated due to heat exchanger problems, which were associated with system plugging or fouling. After completion of each experiment, changes in system control and/or cooling water chemical additions were made. A brief description of the system operational conditions during each of the nine experiments is listed in Table VI-1. Descriptions of the chemical additions used in the experiments are given in Table VI-2.

TABLE VI-2

COOLING WATER CHEMICAL TREATMENT DESCRIPTIONS

Chromate Treatment A

8-10 mg/l; Nalco 364, zinc chromate corrosion inhibitor
HCl for pH Control (pH range 6.5 - 7.0)
10-20 mg/l; Nalco 207, Methylene Bis Thiocyanate

Polyol Treatment B

5-10 mg/l free PO_4; Nalco 345, polyol ester scale inhibitor
2-3 mg/l; Nalco 344, organic dispersant
HCl for pH Control (pH range 6.5 - 7.5)
10-20 mg/l; Nalco 207, Methylene Bis Thiocyanate

Chromate Treatment C

10-12 mg/l; Nalco 37, chromate corrosion inhibitor (free of zinc)
HCl for pH Control (pH range 6.5 - 7.5)
and intermittant use of
50 mg/l, Nalco 207, Methylene Bis Thiocyanate, and Nalco 322, a chlorinated phenol with an amine base

Chromate Treatment D

10-12 mg/l; Nalco 37, chromate corrosion inhibitor (free of zinc)
H_2SO_4 for pH Control (pH range 6.7 - 7.2)
Ozonation of tertiary effluent and sodium hypochlorite for disinfection

Chromate Treatment E

10-12 mg/l; Nalco 37, chromate corrosion inhibitor (free of zinc)
HCl for pH Control (automatic control, set point pH = 7.2)
1/2 pint/day; Nalco 7312, organic dispersant
Sodium hypochlorite only for disinfection

Chromate Treatment F

10-12 mg/l; Nalco 37, chromate corrosion inhibitor (free of zinc)
0.5-1.0 mg/l residual chlorine by gas chlorination of tertiary effluent
HCl for pH Control (automatic control, set point at pH of 6.7 - 6.8)
1/2 pint/day; Nalco 7312, organic dispersant (fed over 24 hours)

The initial pilot trials did little to demonstrate the economic and technical feasibility of using reclaimed industrial wastewater as make-up to the process cooling systems. The continual failures of heat exchangers during these trials demonstrated the need for greater control of microbial growth in the cooling system and greater control of scale deposition in the heat exchangers of the cooling system. An evaluation of the initial trials indicated that the "best" system performance occurred during the use of Treatment C (see Table VI-2). Although operations using this treatment were of limited duration, significant improvements in performance were observed.

The first two experiments of the continuation trials (No. 7 and 8) utilized a chemical treatment scheme similar to that used in experiments No. 5 and 6. Microbial growth in the pilot cooling loop was controlled more adequately during these experiments, primarily through continuous disinfection of the pilot system makeup water. Additionally, the source of microbial food was reduced through the implementation of the improvements to the existing treatment facilities stated in Chapter IV. A reliable chemical treatment and control scheme (Treatment F, Table VI-2), was developed during the end of experiment No. 8. This scheme was then verified in experiment No. 9, the 90-day trial.

Plant production personnel determined that the three month period was the minimum required period of continuous operation of the pilot cooling system needed to evaluate the feasibility of the system.

Average and range concentrations of "key" parameters in the makeup and pilot cooling loop water during the 90-day trial are listed in Table VI-3.

TABLE VI-3

WATER QUALITY DURING 90-DAY TRIAL

Parameter	Makeup Average (mg/l)	Makeup Range (mg/l)	Pilot Cooling Loop Average (mg/l)	Pilot Cooling Loop Range (mg/l)
TOC	28	14-39	150	28-250
Total Dissolved Solids (TDS)	280	175-390	1750	400-2150
Calcium Hardness	35	20-46	185	40-264
Silica	19	10-49	70	22-104
Total Hardness	41	32-54	238	160-280
Sulfate	74	31-159	316	177-414
Zinc	2.0	1.1-6.1	19.5	10.8-32.5

Scale deposition in the heat exchangers of the pilot system was not completely eliminated during the 90-day trial. However, the scale which did form deposited almost immediately and had no effect on heat transfer from the heat exchanger to the cooling water. Based upon these results, the water quality criteria listed below were selected for use in full-scale application of reclaimed industrial wastewater in the plant cooling systems.

TABLE VI-4

COOLING SYSTEM WATER QUALITY CRITERIA*

Parameter	Process Cooling Systems	Condenser Cooling Systems and Chillers
Total Hardness	350 mg/l	450 mg/l
Calcium Hardness	300 mg/l	400 mg/l
Silica	200 mg/l	200 mg/l
Sulfate	500 mg/l	600 mg/l
Zinc	42 mg/l	42 mg/l
pH range	6.6 - 7.2	6.6 - 7.2

*Concentration limits

CHAPTER VII

CONCEPTUAL DESIGN OF ADVANCED WASTEWATER TREATMENT PROCESSES AND INTEGRATED RECIRCULATION PLAN

Preliminary engineering efforts have culminated in the preparation of conceptual designs for tertiary treatment processes. This chapter presents those designs along with an overview of the recirculation scheme for the Anderson Plant. The recirculation plan is indeed a demonstration project, and accordingly not all questions have been answered. However, a rational plan of attack has been defined for further evaluations during final design work and economic analysis.

CONCEPTUAL DESIGN OF ADVANCED WASTEWATER TREATMENT PROCESSES

Treatability studies have produced a wealth of data. Conceptual designs developed here are based not only upon bench and pilot scale performances, but also upon our industrial waste experiences. Sand filtration, activated carbon adsorption, and disinfection processes will be required to produce a reclaimed wastewater relatively free of suspended solids and organic materials. It is possible that some form of inorganic contaminant removal other than drift loss may also be provided.

Sand Filtration

The main function of these filters will be to protect the carbon adsorbers from inordinate solids loadings occurring during secondary clarifier upsets. Either dual-media or multi-media downflow pressure filters seem to be best suited in this application.

Filters

Basic design criteria are as follows:

Filtration Rate = 4 gpm/ft^2
Solids Loading at Breakthrough = 0.10 lb TSS/ft^3
Bed Depth = 36"
Backwash Cycle Length = 20 minutes

Influent TSS Concentration = 9 mg/1

During each filter cycle, the volume of wastewater processed will be:

$$\frac{(0.10\ \text{lb/ft}^3)(3.0\ \text{ft})(10^6\ \text{gal/MG})}{(9\ \text{mg/1})(8.34\ \text{lb/MG/mg/1})} = 3997\ \text{gal/ft}^2$$

Therefore, the filter cycle length will be:

$$\frac{3997\ \text{gal/ft}^2}{4\ \text{gpm/ft}^2} = 999\ \text{minutes}$$

Now, the complete cycle length is:

Complete Cycle Length = Filter Cycle Length + Backwash Cycle Length
= 999 min. + 20 min.
= 1019 minutes

During each day, one may expect

$\frac{1440}{1019}$ cycles or 1.413 cycles/day

Also, the filter time per day will be:

1.413 (999 min) = 1412 minutes

The backwash time will be:

1.413 (20 min) = 28 minutes

Thus, during a 24 hour period, the filters will operate 1412/1440 or 98% of the time. Based upon the design flow rate of 285 gpm,

Effective Flow $= \frac{285}{0.98}$ gpm = 291 gpm

Required Filter Area $= \frac{291\ \text{gpm}}{4\ \text{gpm/ft}^2} = 73\ \text{ft}^2$

If 6' diameter filters are used,

$\frac{73\ \text{ft}^2}{\pi(3\ \text{ft})^2} = 2.6$ will be required

If 7' diameter filters are used,

$\frac{73\ \text{ft}^2}{\pi(3.5\ \text{ft})^2} = 1.9$ will be required

Therefore, use 2 @ 7' diameter downflow pressure filters. The bed expansion during backwash will be approximately 50%; the minimum filter height, not including interior appurtenances, is 1.50 (3.0 ft) = 4.5 ft.

Backwash

Secondary clarifier effluent will be utilized as backwash water in conjunction with an air scour. The major portion of the backwash flow will be introduced into the bottom of the filter, with the remainder to be used as a surface wash. Backwash criteria are listed below.

Backwash Rate = 20 gpm/ft^2
Air Scour Rate = 5 cfm/ft^2

Now,

Backwash Flow = (20 gpm/ft^2)(38.5 ft^2) = 770 gpm
Backwash Volume = (2)(770 gpm)(28 minutes/day) = 43,120 gal/day
Blower Capacity = (5 cfm/ft^2)(38.5 ft^2) = 193 cfm

Backwash effluent will be routed to the Equalization Basins.

Pumps

At least two pumps will be required, one for filter feed and the other for backwash:

Feed Pump:	Vertical, radial flow, 400 gpm (∼10 HP)
Backwash Pump:	Vertical, radial flow, 800 gpm (∼15 HP)

It is advisable to include a spare pump for each purpose.

Filter Feed and Backwash Sump

The filter feed and backwash sump will provide a constant supply of feed and sufficient waters for backwash:

Sump Volume = Daily Backwash Volume
= 43,120 gal or 5765 ft^3
Dimensions: 21' x 25' x 12' deep

Activated Carbon Adsorption

Adsorbers

Carbon adsorption can be provided either by installation of an OCF-owned system or on a service contract basis. The conceptual design presented here has been developed as a general framework for this unit process.

Downflow pressure adsorbers, in series, were tested during treatability studies and are chosen for use. Three adsorbers will be provided, with two to be operating at any time. When carbon in the first adsorber is exhausted, the second adsorber will become the first adsorber and the third will become the second. Assuming one (1) hour per day of downtime,

$$\text{Effective Flow} = 285 \text{ gpm } \left(\frac{24}{23}\right) = 297 \text{ gpm}$$

Now, using a superficial velocity of 4 gpm/ft^2:

$$\text{Surface Area} = \frac{297 \text{ gpm}}{4 \text{ gpm/ft}^2} = 74.25 \text{ ft}^2$$

Therefore, use 10' diameter columns (area = 78.5 ft^2).

Based upon a carbon contact time of 30 minutes, for two filters in series the bed depth is:

$$\text{Total Bed Depth} = \frac{(297 \text{ gpm})(30 \text{ min.})/(7.48 \text{ gal/ft}^3)}{78.5 \text{ ft}^2} = 15.2 \text{ ft}$$

and, Column Bed Depth = 7.6 ft.

The column bed volume is:

$$\text{Bed Volume} = (78.5 \text{ ft}^2)(7.6 \text{ ft}) = 596.6 \text{ ft}^3$$

For 8 x 30 mesh granular carbon, the density is approximately 30 lb/ft^3. Thus, the bed weight is:

$$(596.6 \text{ ft}^3)(30 \text{ lb/ft}^3) = 17{,}898 \text{ lb}$$

It follows that:

$$\text{Adsorber Carbon Inventory} = (17{,}898 \text{ lb})(3) = 53{,}694 \text{ lb}$$

Using a backwash bed expansion of 50%, the expanded bed height

$$= 7.6 \text{ ft} + 0.50(7.6 \text{ ft}) = 11.4 \text{ ft.}$$

Carbon capacity for Anderson secondary effluent is approximately 0.08 lb TOC/1b carbon. Based upon a TOC loading of 116 lb/day and 50% removal, the carbon exhaustion rate is:

$$\text{Carbon Exhaustion Rate} = \frac{(116 \text{ lb TOC/day})(0.50)}{0.08 \text{ lb TOC/1b carbon}}$$

$$= 725 \text{ lb carbon/day (or } 264{,}625 \text{ lb/yr)}$$

Based upon a 10% attrition rate, the annual carbon loss is:

(725 lb/day) (365 days/yr) (0.10) = 26,463 lb/yr

Carbon regeneration will be provided most economically under a contract basis.

Backwash

Backwash criteria are listed below:

Backwash Rate = 15 gpm/ft^2

Backwash Time = 20 minutes

Air Scour Rate = 5 cfm/ft^2

Backwash Water = Carbon Effluent

Then,

Backwash Flow = $(15\ \text{gpm/ft}^2)(78.5\ \text{ft}^2) = 1178\ \text{gpm}$

Backwash Volume = (1178 gpm)(2)(20 min) = 47,120 gal/day

Blower Capacity = $(5\ \text{cfm/ft}^2)(78.5\ \text{ft}^2) = 393\ \text{cfm}$

Backwash effluent will be routed to the Equalization Basins.

Pumps

Two pumps will be required, one for adsorber feed and one for backwash:

Feed Pump: Vertical, radial flow, 400 gpm (10 HP)

Backwash Pump: Vertical, radial flow, 1200 gpm (20 HP)

It is advisable to include a spare pump for each purpose.

Carbon Storage and Transport

Storage must be provided for virgin, regenerated, and spent carbon.

Virgin Carbon Storage = 596.6 ft^3

Spent Carbon Storage = 596.6 ft^3

Use 2 @ 640 ft^3 tanks with 45° hopper bottoms (10' diameter, 11.5' high)

Regenerated Carbon Storage = 596.6 ft^3

Use the spare adsorber.

During carbon transport, at least one gallon of water will be required for each pound of carbon:

$$\text{Water} = \left(\frac{1\ \text{gal}}{\text{lb}} \times 17{,}898\ \text{lb/transport}\right) + 25\%\ \text{contingency} = 22{,}372\ \text{gal/transport}$$

Adsorber Feed Sump

The adsorber feed sump will serve to equalize flow from the sand filters and store feed during adsorber backwash.

Sump Volume = (297 gpm)(120 min)

= 35,640 gal

= 4,765 ft^3

Dimensions: 18' x 25' x 12' deep (Common wall construction with
filter feed sump)

Adsorber Backwash and Effluent Sump

This sump will store carbon-treated effluent for use as backwash water.

Sump Volume = 47,120 gal
= 6300 ft^3

Dimensions: 23' x 25' x 12' deep (Again, common wall construction)

Disinfection

Several disinfecting agents were used during the pilot cooling loop trials, and all, with the exception of chlorine gas, resulted in precipitation of $Fe(OH)_3$. Chlorine will be used here for disinfection, but ferric chloride addition to the primary clarifiers will have to be closely monitored, because phenomena not observed in pilot tests may become apparent in the full scale system. While conventional design dictates a 30 minute chlorine contact time, intermixing of the chlorine gas with the carbon effluent will occur in a flash mix chamber, and the required contact period will be satisfied in the Distribution Tank (old aerobic digester).

Flash Mix Chamber

Based upon a hydraulic retention time of one minute, the required volume is:

Volume = 285 gpm (1 minute)
= 285 gal
= 38 ft^3
Dimensions: = 3.5' x 3.5' x 4'

At a power level of 3 water HP/1000 gal, the required mixer horsepower is:

$$\left(\frac{3 \text{ water HP}}{1000 \text{ gal}}\right) (285 \text{ gal}) \left(\frac{1}{0.70}\right)$$
$$= 1.2 \text{ BHP or } 2 \text{ BHP}$$

Chlorine Dosage

Treatability studies indicated that a chlorine dosage of 2 mg/l will provide nearly complete elimination of fecal coliforms.

Dosage $= 1\text{-}5 \text{ mg/l } Cl_2$

Minimum Feed $= (1 \text{ mg/l})(285 \text{ gal/min})(1440 \text{ min/day})$

$(\bar{M}G/10^6 \text{ gal})(8.34 \text{ lb}/\bar{M}G/\text{mg/l})$

$= 3.4 \text{ lb } Cl_2/\text{day}$

Maximum Feed $= (3.4 \text{ lb } Cl_2/\text{day})(5)$

$= 17 \text{ lb } Cl_2/\text{day}$

Distribution Tank

The old aerobic digester (volume = 297,000 gal) will receive reclaimed wastewater from the flash mix chamber. City water will also enter this tank; thus, the tank will serve as the distribution point for all process and cooling water uses in the plant. Excess wastewater flows will flow to the Storage Basin for use at a later time. Additionally, flows may be returned from the basin to the tank. The tank should be covered to prevent contamination.

Reclaimed Wastewater Storage Basin

This basin will receive excess reclaimed wastewater flows during winter operations.

Basin Volume $= (10 \text{ gpm})(1440 \text{ min/day})(100 \text{ day})$

$\simeq 1.5 \ \bar{M}G$

$= 200.535 \text{ ft}^3$

Flows from the basin will be routed to the Distribution Tank, to the Filter Feed and Backwash Sump, or to the Off-Specification Basin. The existing effluent retention pond will be enlarged into two basins, one for reclaimed wastewater storage and the other for off-specification wastewater storage.

Off-Specification Basin

This basin will receive secondary clarifier effluent during plant upsets. A five-day hydraulic retention time is provided (based upon 205 gpm).

Basin Volume $= (205 \text{ gpm})(1440 \text{ min/day})(5 \text{ day})$

$= 1.5\ \bar{M}G$

$= 200,\ 535\ ft^3$

Water in the basin will be bled into the Equalization Basins at a controlled rate. Backwash effluents from the sand filters and carbon adsorbers (∿63 gpm total) will flow to the Equalization Basins for purposes of flow equalization.

The existing effluent retention pond possesses a volume of approximately 96,000 ft^3; thus, it will have to be more than quadrupled in size. Both basins (storage and off-specification) will be earthen with protected side slopes. Soil in the area consists of relatively impervious clay.

Summary

A process flow diagram for the advanced wastewater treatment facilities is shown as Figure VII-1. Secondary clarifier effluent will flow either into the Off-Specification Basin (upset conditions) or the Filter Feed and Backwash Sump. Also, return flows from the Reclaimed Wastewater Storage Basin may be returned to the sump for treatment prior to use. Wastewater will enter the downflow, pressure sand filters, with filter effluent flowing into the Adsorber Feed Sump. Filter Backwash water will be pumped from the sump up through the filters and into the Equalization Basins. The backwash may thus be returned to the treatment system at a reasonably constant rate, along with "off-spec" return flows.

The Adsorber Feed Sump will equalize the wastewater flow rate prior to feed to the downflow, pressure carbon adsorbers (series configuration). Carbon-treated effluent will flow into the Adsorber Backwash and Effluent Sump, which will store water for use as backwash water. The backwash effluent will be pumped to the Equalization Basins for gradual return to the treatment system.

Chlorine gas will be added in the Flash Mix Chamber, followed by a continued contact period as the reclaimed wastewater flows to and through the Distribution Tank. City water will be added to the Distribution Tank at a controlled rate. The mixture will then be distributed to the process and cooling uses throughout the plant. Alternatively, during periods of extreme/excess flows, the waters may flow into the Reclaimed Wastewater Storage Basin.

This tertiary treatment system has been designed to provide maximum flexibility during operations. The effluent quality from the filtration, carbon adsorption, and disinfection treatment sequence is estimated as follows:

Flow = 285 gpm
TSS = 0 - 5 mg/l
TOC = 10 - 20 mg/l
BOD_5 = 0 - 5 mg/l

INORGANIC CONTAMINANT REMOVAL

The projected daily mass inputs and reclaimed wastewater equilibrium concentrations for total and calcium hardness, silica, sulfate, and zinc given in Table III-11 were calculated using drift loss as the sole removal mechanism. Water quality criteria for the cooling systems (Table VI-4) based upon the results of the pilot cooling loop experiments are detailed in Table VII-1 for each cooling system. Due to the concentrating effect in each evaporative cooling system and in the cascade pattern, the allowable concentration of each constituent in the makeup to a particular system will be considerably less than that in the system itself. Based upon the cycles of concentration for each cooling system and the cascade pattern, a factor which represents the effective concentration ratio in each system has been developed; these are presented in Table VII-2. The maximum allowable reclaimed wastewater concentration for a particular component is then:

$$\text{Concentration} = \frac{\text{Maximum Allowable Concentration in the Cooling System}}{\text{Factor}}$$

These values have been calculated for each system and are presented in Table VII-3.

Removal Requirements

It is obvious that most of these allowable makeup concentrations are much lower than the equilibrium concentrations projected in Chapter III. Present data indicates that inorganic contaminants may have to be removed from the water system by some mechanism other than drift loss. However, this will not be adequately defined until the recycle system is operated during the demonstration period included in the grant schedule. Using the smallest values for each parameter from Table VII-3, the required removals were estimated as follows:

FIGURE VII-1

PROCESS FLOW DIAGRAM - ADVANCED

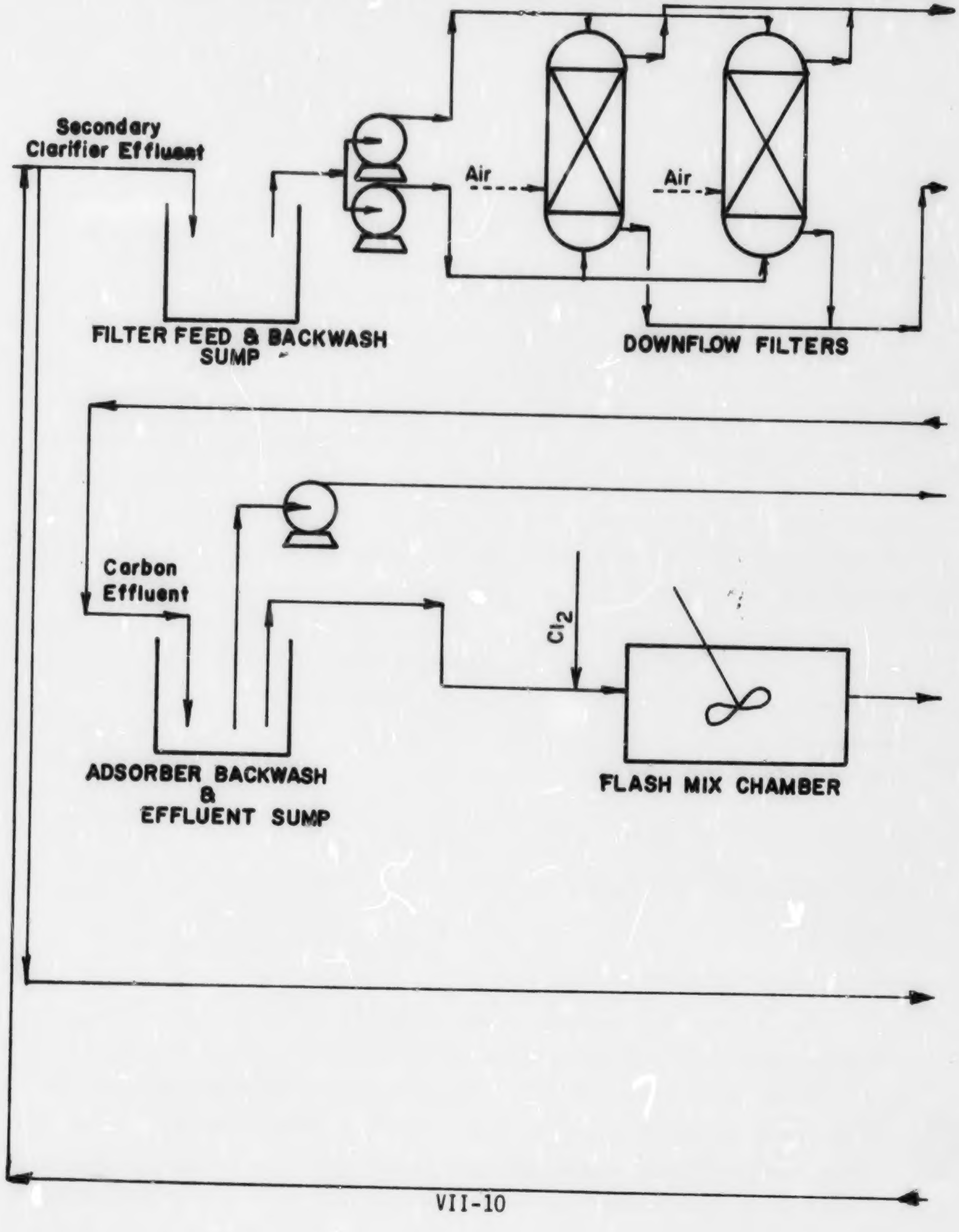

Secondary Clarifier Effluent
Air
Air
FILTER FEED & BACKWASH SUMP
DOWNFLOW FILTERS
Carbon Effluent
Cl_2
ADSORBER BACKWASH & EFFLUENT SUMP
FLASH MIX CHAMBER

FIGURE VII-1

WASTEWATER TREATMENT FACILITIES

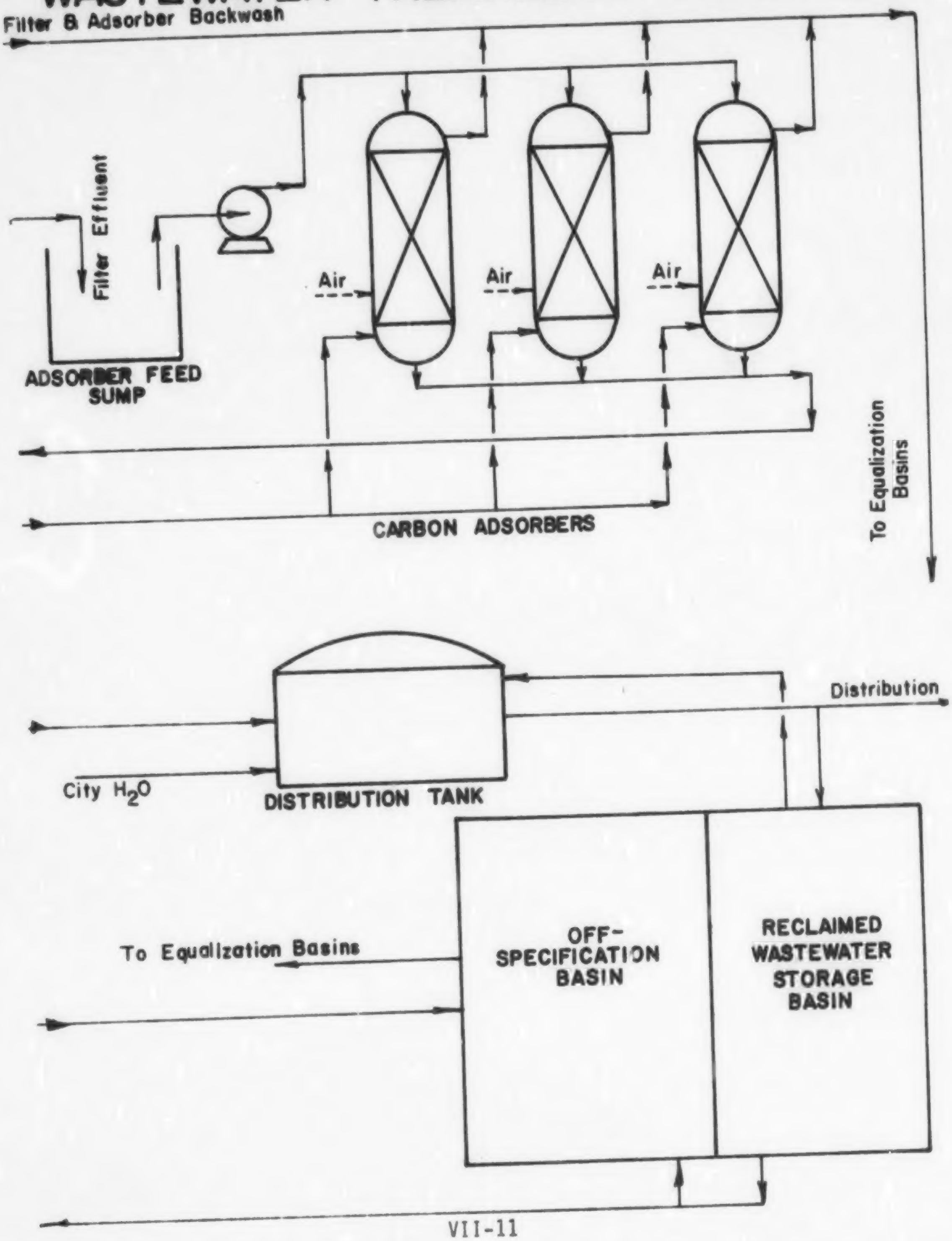


TABLE VII-1

MAXIMUM ALLOWABLE CONCENTRATIONS IN THE COOLING SYSTEMS

System	Total Hardness[1]	Calcium Hardness[2]	Silica[3]	Sulfate	Zinc
"A" Chillers	450	400	200	600	42
"E" Chillers	450	400	200	600	42
"D" Chillers	450	400	200	600	42
#1 Pond	450	400	200	600	42
#2 Pond	350	300	200	500	42
"D" Condenser Cooling	450	400	200	600	42
"D" Process Cooling	350	300	200	500	42
Chemical Cooling Tower No. 1	350	300	200	500	42
Chemical Cooling Tower No. 2	350	300	200	500	42

Notes: 1. mg/l as $CaCO_3$.
2. mg/l as $CaCO_3$.
3. mg/l as SiO_2.

TABLE VII-2

FACTORS USED IN CALCULATION OF ALLOWABLE RECLAIMED WASTEWATER CONCENTRATIONS

System	Summer	Winter
"A" Chillers	1.00	3.50
"E" Chillers	1.00	5.50
"D" Chillers	1.00	7.00
#1 Pond	4.71	4.50
#2 Pond	4.70	8.20
"D" Condenser Cooling	5.00	7.00
"D" Process Cooling	4.57	8.42
Chemical Cooling Tower No. 1	2.33	1.83
Chemical Cooling Tower No. 2	5.83	4.83

TABLE VII-3

MAXIMUM ALLOWABLE RECLAIMED WASTEWATER CONCENTRATIONS

System	Total Hardness		Calcium Hardness		Silica		Sulfate		Zinc	
	S[1]	W[2]	S	W	S	W	S	W	S	W
"A" Chillers	450	129	400	114	200	57	600	171	5	5
"E" Chillers	450	82	400	73	200	36	600	109	5	5
"D" Chillers	450	64	400	57	200	29	600	86	5	5
#1 Pond	96	100	85	89	43	44	127	133	5	5
#2 Pond	74	43	64	37	43	24	106	61	5	5
"D" Condenser Cooling	90	64	90	57	40	29	120	86	5	5
"D" Process Cooling	77	42	66	36	44	24	109	59	5	5
Chemical Cooling Tower No. 1	150	191	129	164	86	109	215	273	5	5
Chemical Cooling Tower No. 2	60	73	51	62	34	41	86	104	5	5
Equilibrium Values	94	95	87	88	42	42	186	188	6	6

Notes: 1. S = summer conditions.

2. W = winter conditions.

Input = Required Removal + Drift Loss

Summer.........Input = $X + 0.840 (W_{min})$

Winter.........Input = $X + 0.831 (W_{min})$

where:

Input = daily mass input of a particular constituent

X = required removal

W_{min} = the most stringent makeup quality required

The concentrations, inputs, drift losses, and required removals are summarized in Table VII-4.

Removal Alternatives

The logical point to remove these contaminants from the water system is where they are highly concentrated, i.e. the cooling system blowdowns. If the blowdowns were mixed with process wastewater, the flow rates would be increased and the concentrations reduced so as to make removal less economical. Based upon removing the inorganics from the system at #1 Pond, #2 Pond, "D" Condenser Cooling, and "D" Process Cooling, several alternative removal schemes have been examined:

- Direct discharge of cooling system blowdown
- Combined discharge of cooling system blowdown with treated sanitary wastewater
- Inorganic salt removal through
  - ion exchange
  - reverse osmosis
  - lime-soda softening and anion exchange

These alternatives are portrayed graphically in Figure VII-2. Each will be discussed in turn with regard to performance and applicability. The discharge or treatment of 30 - 50 gpm of cooling system blowdown is the basis for each alternative mechanism.

Direct Discharge

This alternative is obviously the most straightforward and economical choice. It does fall short of "total" recycle, and concentrations or masses of constituents in the cooling system blowdowns may exceed those allowed. Approximately 30 - 50 gpm of cooling system blowdowns would have to be collected and conveyed to the point of discharge.

TABLE VII-4

INORGANIC REMOVALS

Summer

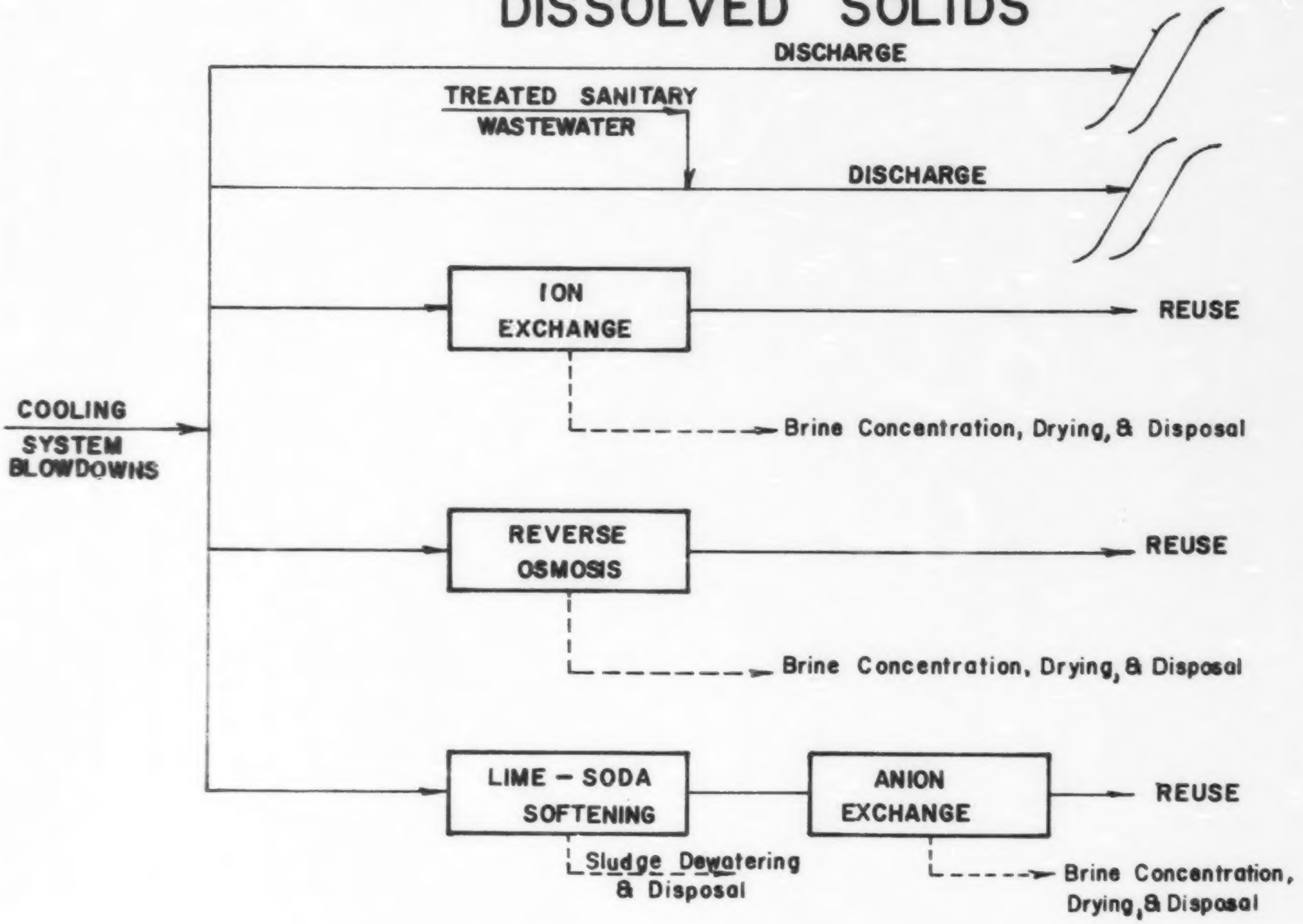
Parameter	Concentration	Input	Drift Loss	Removal Required
Total Hardness	60 mg/l	79 #/day	50.4 #/day	28.6 #/day
Calcium Hardness	51	73	42.8	30.2
Silica	34	35	28.6	6.4
Sulfate	86	156	72.2	83.8
Zinc	5	4.8	4.2	0.6

Winter

Parameter	Concentration	Input	Drift Loss	Removal Required
Total Hardness	42 mg/l	79 #/day	34.9 #/day	44.1 #/day
Calcium Hardness	36	73	29.9	43.1
Silica	24	35	19.9	15.1
Sulfate	59	156	49	107
Zinc	5	4.8	4.2	0.6

FIGURE VII-2

ALTERNATIVES FOR REMOVAL OF INORGANIC DISSOLVED SOLIDS


DISCHARGE
TREATED SANITARY WASTEWATER
DISCHARGE
COOLING SYSTEM BLOWDOWNS
ION EXCHANGE
REUSE
Brine Concentration, Drying, & Disposal
REVERSE OSMOSIS
REUSE
Brine Concentration, Drying, & Disposal
LIME – SODA SOFTENING
Sludge Dewatering & Disposal
ANION EXCHANGE
REUSE
Brine Concentration, Drying, & Disposal

Combined Discharge

This scheme involves discharge of cooling system blowdown along with treated sanitary wastewaters. Again, the blowdowns would have to be collected and conveyed to the point of discharge, and problems could arise with respect to effluent limitations. This alternative, like the previous one, does not allow "total" recycle.

Ion Exchange

Mixed-bed ion exchange tests conducted as part of the treatability studies provided a starting point for evaluation of ion exchange for removal of calcium, magnesium, silica, sulfate, and zinc from cooling system blowdowns. In order to exchange both negative and positive ions and not add significant quantities of dissolved solids to the water, a mixed bed of a strong acid resin and a strong base resin would be utilized. Design calculations revealed that low mixed-bed resin exchange capacity demonstrated during the treatability studies and high regenerant requirements would result in an applied water recovery of 54%. Also, the remaining volume of water would contain appreciable quantities of dissolved solids and presents a disposal problem in itself. Basically, the brine would have to be concentrated through evaporation or freeze crystallization and dryed; the remaining salts could be disposed of in a landfill. In conclusion, ion exchange does not appear to be suited to this application either from the standpoint of cost or that of water recovery.

Reverse Osmosis

Osmosis is the transport of a solvent from a dilute to a concentrated solution across a semipermeable membrane. The transport is caused by a chemical potential driving force manifested as the osmotic pressure. If pressure in excess of the osmotic pressure is applied to the concentrate side of the membrance, the direction of solvent flow is reversed so that solvent flows across the membrance from the concentrated to the dilute phase. This phenomenon is reverse osmosis.

In late 1974 the ROGA Division of Universal Oil Products Co. conducted laboratory reverse osmosis (RO) tests on the Anderson secondary clarifier effluent. The wastewater was neutralized to pH 5.4 and filtered through a 25 micron filter prior to being fed to a spiral wound cellulose acetate RO module at 400 psig. Test results were fairly encouraging, as a water recovery of 86.8% was obtained at a flux of 12 gal/ft^2-day, along with solute rejection factors of 94-97% for conductivity.

Based upon the test results, 30 gpm would require an RO membrane surface area of:

$$\frac{(30 \text{ gpm})(1440 \text{ min/day})}{12 \text{ gal/ft}^2 \text{ day}} \quad \text{or} \quad 3600 \text{ ft}^2$$

A reject stream of 30 gpm (1-0.868), or 4.1 gpm, would have to be concentrated and dryed. The remaining salts could then be landfilled. In summary, the RO process will probably perform quite well in this application, but at considerable costs. Scale formation on membrane surfaces by calcium carbonate and calcium sulfate could dictate chemical pretreatment in additon to pH adjustment and filtration.

Lime-Soda Softening and Anion Exchange

This treatment sequence would involve lime-soda softening for removal of Ca^{++}, Mg^{++}, SiO_2, and Zn^{++}, followed by anion exchange (weak base resin) for $SO_4^=$ removal. Of course, softening will produce a sludge (to be dewatered and disposed of) and anion exchange regenerant streams will have to be concentrated and dryed prior to disposal. Initial process design calculations resulted in the following design criteria:

Lime-Soda Softening

Lime Dose = 131 mg/l as CaO
Lime Feed = 52 lb/day (pulverized quicklime)
Soda Ash Dose = 330 mg/l as Na_2CO_3
Soda Ash Feed = 120 lb/day

Chemical Feeders:

Lime	2.0 lb/hr to 4.0 lb/hr
Soda Ash	5 lb/hr to 10 lb/hr

Solids Contact Clarifier:

10' diameter
7' side water depth

Recarbonation:

Volume = 250 gal
Dimensions: 3' x 3' x5'
CO_2 Dose = 0-100 mg/l (pH 9.4)
CO_2 Feed = 0-18 lb/hr

Rapid Sand Filter:

Area = 20 ft^2

Sludge Generation: 182 lb/day (dry weight)

Anion Exchanger

Resin Capacity = 24.6 eq/ft^3
Exchanger = 5' diameter
10' height
Resin Volume = 90.8 ft^3
Regenerant = 43 gal/day @ 50% NaOH
Regeneration Rate = 45.5 gpm (17.3 min)
Backwash Rate = 20 gpm (5 min)
Rinse Rate = 136.2 gpm (33.3 min)
Effective Feed Rate = 60 gpm
Concentrate Stream = 3.8 gpm

Summary

Several alternative methods for removal of inorganic contaminants have been examined. The two involving discharge are not appropriate to the concept of "total" recirculation and would be considered only as temporary measures. Of the three treatment methods, reverse osmosis or lime-soda softening/anion exchange are most promising. Final selection should be based upon economic analyses. At this time, the scheme of action should be to allow for possible discharge of cooling system blowdowns until removal requirements are better defined by the full scale system, and then to implement the most economical, reliable treatment procedures.

INTEGRATED RECIRCULATION PLAN

Recirculation plans for the Anderson Plant involve:

- Treatment of process wastewaters to a quality level that is suitable for recirculation (in terms of suspended solids and organic materials)

- Segregation and separate treatment of sanitary wastewater
- Reuse of reclaimed wastewater in the process areas and cooling systems
- Removal of inorganic contaminants through drift loss or other mechanisms as required.

Water and wastewater distribution, usage, and treatment plans are shown in Figure VII-3. Numbers on the illustration are projected flow rates in gallons per minute (gpm). Existing wastewater treatment facilities are to be supplemented by sand filtration, activated carbon adsorption, and disinfection. Reclaimed wastewater along with city water will be distributed to the process areas and cooling systems. The cooling systems cascade pattern will result in concentration of inorganic solids which will be removed by drift loss, discharge, or treatment. Finally, sanitary wastewaters will be segregated from process wastewaters and treated in a separate package plant. In conclusion, one should realize that this plan is a demonstration project, and as such, not all problems have been answered or even defined. Actual operating experiences will be the ultimate "test" for this recirculation system.

FIGURE VII-3

ANDERSON PLANT WATER

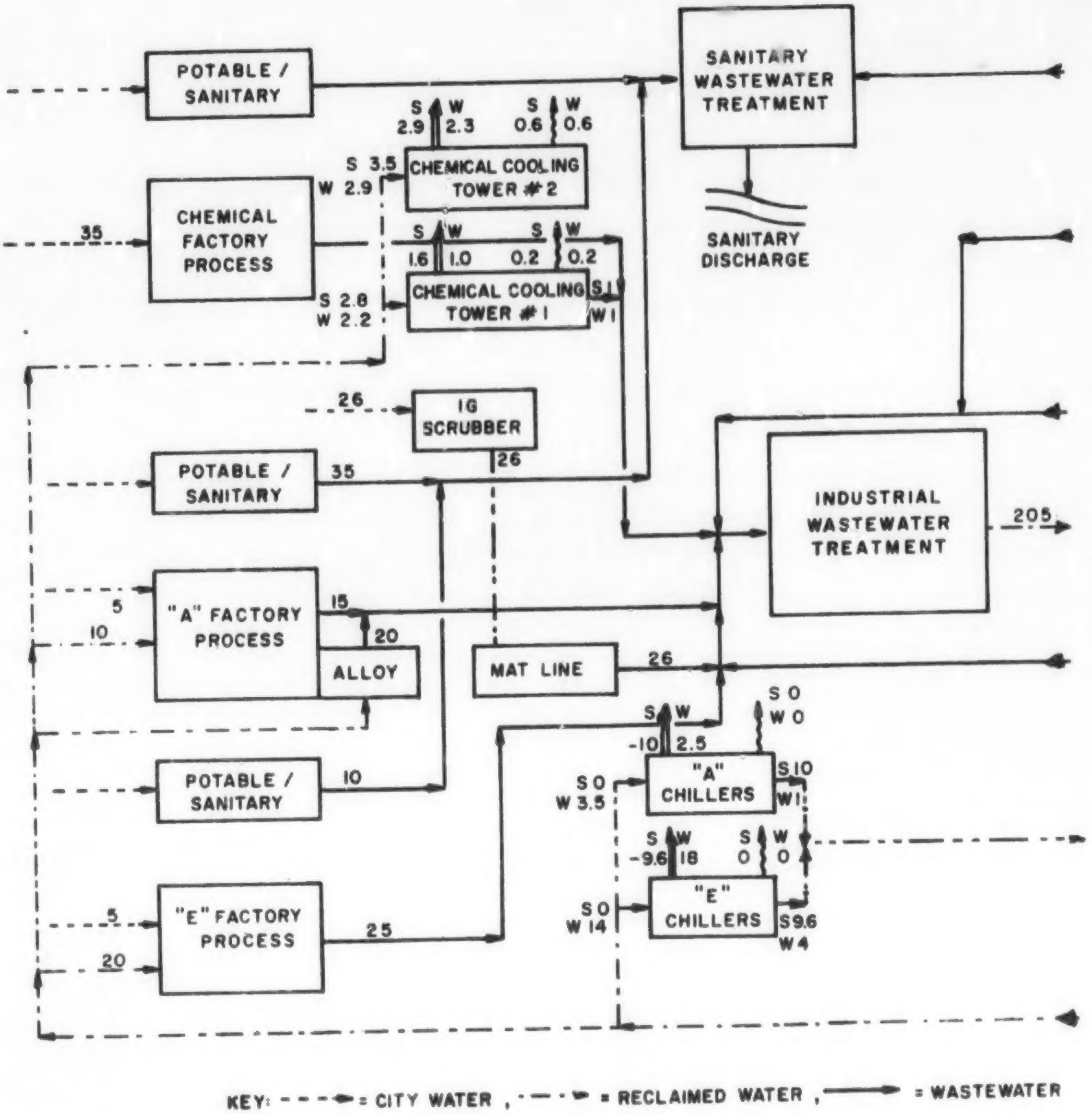


FIGURE VII-3

DISTRIBUTION AND USAGE